

Pilot Scale Trials on Atmospheric Fluidized Bed Combustion of High-Sulphur Eastern Canadian Coals (Minto and Devco) with Limestone Addition for Sulphur Capture

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H.A. Becker and R.K. Code
Department of Chemical Engineering
Queen's University at Kingston
Kingston, Canada

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Quebec Fluidized Bed Combustion Laboratory

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**PILOT SCALE TRIALS ON ATMOSPHERIC FLUIDIZED BED COMBUSTION
OF HIGH-SULPHUR EASTERN CANADIAN COALS (MINTO AND DEVCO)
WITH LIMESTONE ADDITION FOR SULPHUR CAPTURE**

**H.A. Becker and R.K. Code
Department of Chemical Engineering
Queen's University at Kingston
Kingston, Canada K7L 3N6**

**A report to
Energy Mines and Resources Canada
Energy Research Laboratories, CANMET
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ABSTRACT

Combustion of high-sulphur Eastern Canadian coals (Minto and Devco, respectively 7.4 % and 5.4 % S) with addition of an Eastern Canadian limestone (Havelock) for sulphur capture has been studied in a pilot-scale atmospheric fluidized bed combustion (AFBC) system. The combustion chamber is of square cross-section, 0.155 m² in area, and 3.3 m high. Operating conditions were: mean coal particle size 6.6--19.3 mm, mean limestone particle size 0.49--1.98 mm, coal feedrate 26--50 kg/m² hr, Ca/S molar ratio 1.4--4.5, excess air level - 20--44 %, bed temperature 800--960 °C, fluidizing velocity 1.6--2.9 m/s, and cyclone dust recycle 0--100 %. Both coals burned well. Carbon burnups were 88.5--97.5 % of the fuel carbon. An empirical correlation relates carbon loss to fluidizing velocity, limestone particle diameter, excess air level, bed temperature, and recycle level. Havelock limestone (97 % CaCO₃) is a fairly good SO₂ sorbent; 90 % sulphur capture is achieved with a Ca/S ratio of 2.8 using stone of mean particle size under 1 mm, and 98 % at a ratio of 3.6. Only 9--24 % of the fuel nitrogen appears as NO_x, the rest being presumably converted to N₂. The partition of the solid products between bed product, cyclone product (dust not recycled), and baghouse product is determined principally by the recycle level. Other results presented include: heat transfer coefficients to in-bed tubes, freeboard burning and freeboard NO_x reduction, phenomena and characteristics of transient response and statistically stationary fluctuations, mineral element balances (Ca, Mg, Fe, Si, Na, K, V), and fluidization parameters.

PREFACE

This is the Final Report on work performed under a contract with Supply and Services Canada:

DSS Contract Serial No.: OSU81-00226

Title: Commissioning of a pilot scale fluidized bed combustion research apparatus and conducting of a series of tests to establish a standard test procedure to assess the performance of Canadian limestone and coals in fluid bed combustion

Principal investigators: Dr. H.A. Becker and Dr. R.K. Code (Dept. of Chemical Engineering) and Dr. J.T.N. Atkinson (Dept. of Metallurgical Engineering), Queen's University at Kingston, Kingston, Ontario K7L 3N6. Telephone (613) 547-3045

Scientific Authority: Dr. E.J. Anthony, Building 1, Energy Research Laboratories, CANMET, Haanel Drive, Bells Corners, Ontario.
Postal address: Energy Research Laboratories, CANMET, 555 Booth Street, Ottawa, Ontario K1A 0G1. Telephone (613) 996-4570, ext. 185

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Science Procurement Manager: Ms. Sharon McMullin, Science Procurement Branch, Supply and Services Canada, Place du Portage, Phase III, 11 Laurier Street, Hull, Quebec K1A 0S5

Two professionals were fully employed in work on the contract from its beginning: Mr. David Liang (Engineer) and Mr. John Stephenson (Technician). A third, Mr. Robert McCleave (Engineer) joined in the second year. Two Ph.D. students doing their thesis research on fluidized bed combustion participated: Mr. Michel Couturier and Mr. Fernando Preto. A Visiting Scholar from the People's Republic of China, Bing-Song Xu, was fully involved during his year in Canada. Two undergraduate students were employed each summer as Summer Research Assistants: Mr. Steven Butler and Mr. Donald Swain in 1981, and Mr. Bruce Goddard and Miss Marilyn McInerney in 1982.

A number of bench-scale and theoretical studies have been carried on complementing the pilot plant work. A part of the costs of these, and all of the costs of supporting graduate students, the second engineer, and some of the summer research assistants, were met through a Strategic Energy Grant from the Natural Sciences and Engineering Research Council of Canada.

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NOTATION

N.1 General remarks

The AFBC pilot plant operation is complex enough so that, as in the treatment of large plants generally, a systematic approach to notation is desirable and even necessary. In the following summary, the system adopted in this report is first explained. The list of symbols and definitions is then presented.

N.2 Intensive variables

The symbols employed for the major intensive variables are ones quite commonly used in the related theory of transport phenomena, thermodynamics, and chemical reactor analysis: T for temperature, P for pressure, ρ for density, c_p for specific heat capacity at constant pressure, w_i for the mass fraction of component i (element or compound) in a material, and x_i for the mole fraction of species i in a molecular mixture. In the specification of material composition, the concepts of mass ratio W_i and mole ratio X_i are also used but (unless otherwise noted) only in the following sense: W_i is the mass of component i per unit mass of water-free (moisture-free) material, and X_i similarly is the molar mass of i per unit molar mass of water-free molecular mixture. Thus

$$W_i = w_i / (1 - w_{H_2O}),$$

$$X_i = x_i / (1 - x_{H_2O}).$$

These concepts are useful for discussing solids compositions on the dry basis, and gas compositions on a water-free ("dry gas") basis. Our gas analysis instruments, for example, were perforce set up to give dry-gas analyses.

N.3 Extensive variables

Standard symbols are employed herein for these too: m for

inertial (ordinary) mass of material, n for molar mass of material (moles), V for volume of material, Q for heat, etc.

N.4 Fluxes

Instead of introducing wholly new symbols for fluxes (mass material flux, molar material flux, heat flux, etc.), we follow the convenient and economical convention of using the symbol for the extensive variable with a dot placed over it. Thus $\dot{m}$ is mass material flux, $\dot{n}$ is molar material flux, $\dot{V}$ is volumetric material flux, $\dot{Q}$ is heat flux, etc. This fluxional dot notation should not be confused with the now generally disused Newton dot notation for the time derivative; e.g., $\dot{m}$ is not dm/dt , the time-rate of change of mass of a system, but simply mass flux (the mass flowrate of a material stream).

N.5 Stream labelling

For a simple system, single-letter subscripts may be adequate to identify all material streams, and possibly none may be needed. Here, however, a more complex but systematic approach is warranted which can easily handle any present or future problems. In general, a subscript m denotes the m 'th stream in a set. Streams are placed in three categories: feed streams (entering the plant), product streams (leaving the plant), and transfer streams (between units in the plant). We then label specific feed streams as $F_1, F_2, F_3, \dots$; specific product streams as $P_1, P_2, P_3, \dots$; and specific transfer streams as $T_1, T_2, T_3, \dots$.

Figure N.1 shows the basic streams in the present pilot plant and their assigned labels. Table N.1 explains the labels. Some special notational questions are dealt with as follows:

- o the pilot plant has numbers of two-phase streams, generally composed of co- or counter-flowing solids and gas. Suppose it is wished to consider only the solids component of stream T_6 ; we then append a symbol S (for solids), making it T_6S .

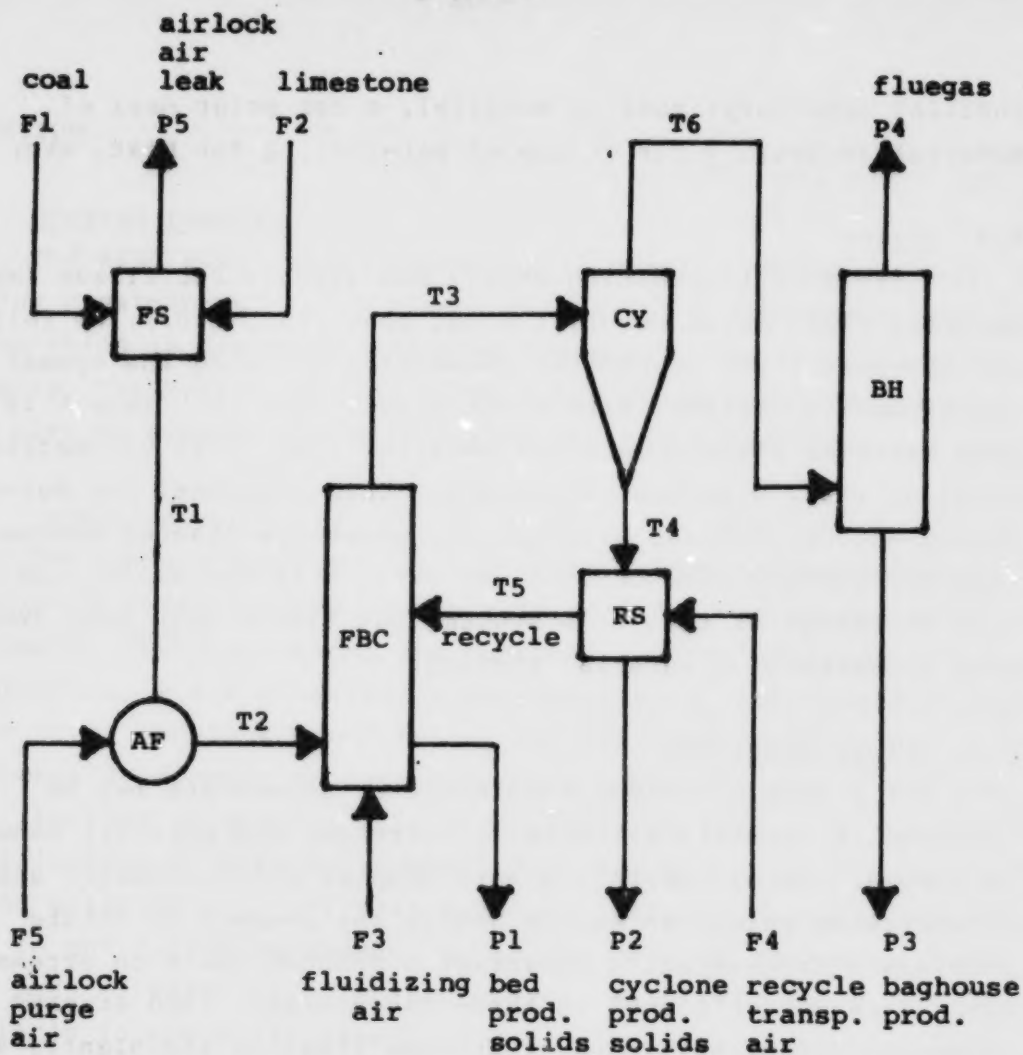


Fig. N.1. Flowsheet of the pilot plant showing the material streams. See Table N.1 for detailed descriptions of the streams. System components:

AF	solids-feed (coal and stone)
	airlock seal and injection system
BH	baghouse
CY	cyclone separator
FBC	fluidized bed combustor
FS	coal and limestone feed system
RS	recycle system

If we were considering only the gas, it would be T6G.

- o it is necessary in some contexts to consider a combination or set of streams, e.g., all the solid products leaving the system, or all the air supplied to the combustor. In the general case, such sets might be labelled like individual streams; e.g., F13 might identify a certain set of feed streams, and F13S the solids component of that set. However, because the number of such cases actually considered herein is small, we have opted for distinctive alphabetical labels for the important ones, Table N.1.
- o for certain basic feedstreams and product streams which are common to all FBC processes and which are involved in many arguments herein, we have defined alternative single-letter labels, Tables N.1 and N.2, which are used almost everywhere in the text.

N.6 Flux ratios

The symbol ϕ_{mn} is used to denote the mass flux ratio $\dot{m}_m / \dot{m}_n$, where m denotes the m'th material stream and n the n'th material stream. Both m and n can represent combinations of, as well as single, streams. This concept is particularly useful in showing how fluxes are partitioned; e.g.:

- o $\phi_{mp} = \dot{m}_m / \dot{m}_p$, m = b, c or h, gives the partitioning of solid products between bed product, cyclone product, and baghouse product
- o $\phi_{C,mf} = \dot{m}_{C,m} / \dot{m}_{C,f}$, m = b, c or h, gives the carbon loss in the bed product, cyclone product, and baghouse product as a fraction of the fuel carbon fed.

N.7 Summary of notation

Table N.2 lists all the symbols used and their definitions.

Table N.1. Notation for identifying material streams. Read in conjunction with Fig. N.1.

General Code	Special short code	Stream
Feed streams:		
F1	f	fuel (here coal)
F2	s	sorbent (here limestone)
F3	--	fluidizing air
F4	--	recycle transport air
F5	--	air-lock purge air
Product streams:		
P1	--	bed product (here bed material removed as overflow)
P2	--	cyclone product (cycloned dust removed as product)
P3	--	baghouse product (dust filtered out by baghouse)
P4	--	stack product, two-phase, consisting of gaseous products and particulate emissions; the "fluegas"
P5	--	leak of air-lock purge air to atmosphere, from the air-lock valve
Transfer streams:		
T1	--	solids feed system to air-lock valve, two-phase, coal and limestone going forward and leaking air-lock purge air going backward (the only counterflow two-phase stream)
T2	--	air-lock valve to combustor, two-phase, consisting of coal, limestone, and air-lock purge air

Table N.1 cont'd

T3	--	combustor freeboard to cyclone, two-phase, consisting of gaseous products and elutriated solids
T4	--	cyclone to recycle system, two-phase, consisting of cycloned dust and leaking exhaust gas
T5	--	recycle system to combustor, two-phase, consisting of recycling dust, recycle transport air, and recycling leaked exhaust gas
T6	--	cyclone to baghouse, two-phase, consisting of gaseous products and of dust escaping the cyclone

Components of two-phase streams:

P4G	--	gaseous portion of the "fluegas"
P4S	--	solid portion of the "fluegas", the particulate emissions
T1S, T2S	--	solid portion of streams T1 and T2, the coal and limestone going to the combustor
T1G	--	purge air leaking from air lock valve, feeding product stream P5
T3S	e	elutriated solids leaving combustor, going to the cyclone
T3G	x	gas leaving combustor, going to cyclone (code x refers specifically to values at combustor exit)
T4S	d	cycloned solids (dust) leaving cyclone
T4G	--	gas leak from cyclone, due to induction by dust recycle system
T5S	--	recycling solids, cyclone to combustor
T5G	--	gas carrying the recycling solids; recycle transport air plus gas leak T4G
T6S	--	solids escaping cyclone, to baghouse
T6G	--	gas leaving cyclone for baghouse

Table N.1 cont'd

Combinations of separate streams:

FA	a	= $F3 + F4 + T2G = F3 + F4 + F5 - P5$, the combustion air
PO	o	= $P2 + P3$, the overhead product solids
PS	p	= $P1 + P2 + P3 + P4S$, the product solids (or product particles)

Table N.2. Summary of notation

Subscripts--special codes for material streams (see Table N.1 for a general listing):

a	combustion air, in aggregate
b	bed product (material removed directly from the bed, here by gravity overflow)
c	cyclone product (cycloned solids not recycled)
d	cycloned solids (dust)
e	elutriated solids (flyash)
f	fuel (here coal)
h	baghouse product (solids)
m	the m'th material stream or combination of streams
n	the n'th material stream or combination of streams
o	overhead product solids (cyclone c and baghouse b combined)
p	product solids (particles), in aggregate
r	recycled cycloned solids
s	sorbent (here limestone)
x	gas at combustor exit

Variables:

A_b	cross-sectional area of bed, m^2
B'	heat transfer conductance per unit length of in-bed tube, tube-side fluid to bed, $W/m \cdot K$

Table N.2 cont'd

c_p	specific heat capacity at constant pressure, J/kg K
COT	continuous-mode operating time, s (see Sect. 3.4)
D_i	inside tube diameter, m
D_o	outside tube diameter, m
D_p	particle diameter, m. Here determined by sieve analysis
$\hat{D}_p$	characteristic diameter for particles of mixed size, m: defined by $D_p = D_p$ at $F_D = 1/2$
F_D	distribution function, $F_D = F_D(D_p)$, of particle size in a population: the mass fraction of particles smaller than the argument value of D_p ; the "mass fraction undersize". Here determined by sieve analysis
g	acceleration due to gravity, m/s ²
h_i	tube-side (inside) heat transfer coefficient for in-bed tubes, W/m ² K
h_o	bed-side (outside) heat transfer coefficient for in-bed tubes, W/m ² K
$\Delta h_{c,f}$	specific enthalpy of combustion of fuel, J/kg
H_s	settled bed height, m: the bed height observed after gradually reducing the gas flow until fluidization ceases and the bed comes to rest (becomes "fixed")
k	thermal conductivity, W/m·K
k_w	thermal conductivity of tube wall, W/m·K
K_b	ratio of pressure drop across bed to weight of bed per unit area

Table N.2 cont'd

L	total in-bed length of the tubes in a series-connected cooling circuit, m
m	mass of material (inertial), kg
m_b	bed mass (holdup), kg
$\dot{m}$	mass flux, kg/s
$\dot{m}_m$	mass flux of the m'th material stream, kg/s. See list of subscripts at head of this table for specific codes
$\dot{m}_{D,m}$	size-resolved mass flux of the m'th material stream, kg/m ³ ·hr, so $\dot{m}_{D,m} dD_p$ is the mass flux of particles and is the size range $D_p, D_p + dD_p$
$\dot{m}_{i,m}$	mass flux of component i (compound or element) carried by the m'th material stream, kg/s. Some specific components: CaCO ₃ , CaO, CO ₂ , SO ₂ , C, H, N, S
n	molar mass of material, mol
$\dot{n}$	molar flux of material, mol/s
$\dot{n}_m$	molar flux of the m'th material stream, mol/s
$\dot{n}_{i,m}$	molar flux of molecular species i carried by the m'th material stream, mol/s
n_t	number of in-bed tubes
Nu	$= h_i D_i / k$ Nusselt number for tube-side heat transfer
P	pressure, Pa
P^*	standard pressure, 101.325 kPa
P_{fb}	freeboard gas pressure, Pa

Table N.2 cont'd

$(\Delta P)_b$	pressure drop across bed, Pa
Pr	v/a , Prandtl number (here for coolant in in-bed tubes)
Q	heat, J
$\dot{Q}$	heat flux out of system, W
$\dot{Q}_b$	heat flux out of bed, W
$\dot{Q}'$	heat flux per unit length of in-bed tube, W/m
r	$= \dot{m}_r / \dot{m}_d$, recycle fraction: fraction of cycloned dust that is returned to combustor
R	$= \dot{m}_r / \dot{m}_c$, recycle ratio: ratio of recycled cycloned dust to that not recycled
R	multiple correlation coefficient (in multiple linear regression analysis)
Re	$= 4\dot{m} / \pi \mu D_i$, Reynolds number for coolant flow through in-bed tube
Re_{mf}	$= D_p v_{mf} / \nu$, particle Reynolds number in bed at minimum fluidization
S	surface, m ²
S_p	particle surface, m ²
S'''	surface area of particles in unit volume of space (in bed or in a bulk sample), m ⁻¹
t	time, s
T	temperature, K
T°	standard temperature, 273.15 K
T_b	bed temperature, K

Table N.2 cont'd

T_i	inside surface temperature of in-bed tube, K
T_o	outside surface temperature of in-bed tube, K
T_w	wall temperature of in-bed tube (recognizing that $T_o = T_i$), K
$T_{w,O}$	value of T_w at start of an in-bed cooling circuit, K
$T_{w,L}$	value of T_w at end of an in-bed cooling circuit, K
v	superficial gas velocity up combustor, m/s. Also called the fluidizing velocity. Here computed as $\dot{n}_x V^*(T_b/T^*)(P^*/P_{fb})$, where $\dot{n}_x$ is the molar gas flux at combustor exit.
v_{mf}	minimum fluidizing velocity, m/s: the value of v at the point of incipient fluidization as indicated by the graph of $(-\Delta P)_b$ vs. v , following the standard procedure
v_{ms}	minimum slugging velocity, m/s: value of v at the onset of slugging
$\tilde{V}^*$	standard molar volume, 22.413 m ³ /kmol
V_p	particle volume, m ³
w_i	mass fraction of component i (compound or element) in a material
$w_{i,m}$	value of w_i in the m 'th material stream
W_i	mass ratio of component i (element or compound) in a material. Here defined only as $W_i = w_i/(1 - w_{H_2O})$, thus as the mass fraction d.b. (dry basis)
$W_{i,m}$	value of W_i in the m 'th material stream
x	distance along an in-bed cooling circuit, m

Table N.2 cont'd

x_i	mole fraction of molecular species i in a molecular mixture
$x_{i,m}$	value of x_i in the m 'th material stream
X_i	mole ratio of molecular species i in a molecular mixture. Here defined only as $X_i = x_i / (1 - x_{H_2O})$, thus as the mole fraction d.b. (dry basis)
$X_{i,m}$	value of X_i in the m 'th material stream
α	$= k / \rho c_p$, thermal diffusivity
ϵ	void fraction in a bed (fraction of space not occupied by particles)
ϵ_a	excess air level: $\dot{n}_a / \dot{n}_{a,st} - 1$, where $\dot{n}_{a,st}$ is the stoichiometric (theoretical) air requirement for complete combustion. Also expressed as a percentage
η_C	carbon burnup efficiency, %: $100(1 - \dot{m}_{C,p} / \dot{m}_{C,f})$
η_Q	bed cooling coefficient: $\dot{Q}_b / \dot{m}_f (-\Delta h_{c,f})$
η_{SO_2}	SO_2 capture efficiency, %: $100(1 - \dot{n}_{SO_2,x} / \dot{n}_{S,f})$
μ	viscosity, $kg/m \cdot s$
μ_w	value of coolant viscosity at temperature T_w , $kg/m \cdot s$
ν	$= \mu / \rho$, kinematic viscosity, m^2/s
ρ	density, kg/m^3
ρ_p	particle density, kg/m^3
ρ_g	gas density, kg/m^3

Table N.2 cont'd

τ_b	$= \dot{m}_b / \dot{m}_p$, estimate of residence time of bed product in combustor, s
τ_p	$= \dot{m}_b / \dot{m}_p$, mean residence time of solid products (in aggregate) in combustor, s
ϕ_a	mixture strength: $\dot{m}_a / \dot{m}_{a,st} = 1 + \epsilon_a$, ratio of air supplied to air theoretically (stoichiometrically) required for perfect combustion
ϕ_{mn}	$= \dot{m}_m / \dot{m}_n$, mass flux ratio of the m'th and n'th material streams
$\phi_{i,mn}$	$= \dot{m}_{i,m} / \dot{m}_{i,n}$, mass flux ratio of component i (compound or element) in the m'th and n'th material streams

N.8 Data symbols

One or the other of two standard sets of data symbols is used in many of the figures. The data symbols are explained in Table N.3; the figure captions then refer to that table for the explanation.

Table N.3. Standard data symbols. When these are used in a figure, then this table is referred to for the key.

A. Identification of limestone particle size and coal type:

Limestone particle size $\hat{D}_p$, mm: $\bigcirc$, 0.49; $\square$, 0.78;
 $\triangle$ $\blacktriangle$, 1.72; ∇ $\blacktriangledown$ 1.98

Coal type: $\blacktriangle$, Devco; others, Minto

B. Identification of coal particle size, coal type, and range of fluidizing velocity:

Coal particle diameter $\hat{D}_p$, mm: $\bigcirc$ $\bullet$, 19.3; $\square$ $\blacksquare$, 10.7;
 $\triangle$ $\blacktriangle$ $\times$ $+$, 6.6.

Coal type: $\times$ $+$, Devco; others, Minto.

Range of fluidizing velocity v , m/s: $\bigcirc$ $\square$ $\triangle$ $\times$, 1.6--2.0;
 $\bullet$ $\blacksquare$ $\blacktriangle$ $+$, 2.7--3.0

1. INTRODUCTION

1.1 Background

A pilot plant for studying atmospheric fluidized bed combustion (AFBC) was built under the preceding contract in the period 1979--81 and is described in a report on that activity (Becker and Code 1982). The present program inaugurated the work the pilot plant was intended for: to investigate the burning of Canadian candidate fuels by AFBC, particularly high-sulphur fuels, using appropriate Canadian limestones as sorbents for SO_2 removal. In addition to testing specific fuels and limestones, it was sought to develop and optimize the test procedures themselves, including the use of bench-scale systems where possible to avoid or minimize more costly and time-consuming pilot plant trials.

1.2 The work undertaken and done

1.2.1 Specifications of the pilot plant

At the start of the present contract, some subsystems of the pilot plant were not yet fully operational. The contract thus included a statement summarizing the required capabilities of the finished system:

- o fluidize beds up to 1 m in slumped depth
- o fluidizing velocities of 1--4 m/s
- o bed temperatures of 700--1100 °C
- o fuel feed system handling fuels such as coal and coke up to 25 mm maximum particle size, of varying size consist, and of moisture content up to 50 %, at feedrates up to 100 kg/hr
- o continuous operation up to 24 hr at expanded bed depth of

0.3--1.0 m with a range of bed materials including silica, limestone, coal ash and olivine

- o sorbent feed system handling relatively dry material up to 6 mm maximum particle size, of varying size consist, at feedrates up to 50 kg/hr
- o recycle system allowing continuous recycle of cycloned dust
- o instrumentation system providing continuous or nearly continuous monitoring, data logging, and data processing by means of a dedicated on-line computer. Information logged to include bed and freeboard temperatures and fluegas concentrations of CO₂, CO, O₂, NO_x and SO₂
- o facilities for all necessary preparation of solids prior feeding, including size reduction, screening, mixing and drying, adequate for at least ten one-day tests per month and at least one test of 24 hr duration

The system as completed and operated during the contract period met most of these specifications. The following exceptions are noteworthy:

- o because of the heavy labour, storage, and transportation requirements, it was found necessary to use locally available commercial services for the crushing, screening, and moving of coal and limestone. Our own facilities were used for mixing, finer screening when needed, and drying when needed
- o the PDP 11/10 computer with which we started suffered a catastrophic disk crash and was found to be not worth repairing, because of the extent of the damage and, at 10 years of age, obsolescence. A new computer was ordered to replace it but, because of slow delivery and the considerable time required for its incorporation into the

system to properly perform its tasks, the computer-based monitoring, data logging, and data processing was not fully available until near the end of the contract period. These functions were thus partly carried out by human operators.

1.2.2 Specific tasks and objectives
for the pilot-plant trials

The tasks set out by the contract to be done by means of the pilot plant were:

1. Conduct shakedown combustion trials, using an easily ignited lignite coal of which a stock was on hand, with the objective of:

- (a) Correcting any problems in the performance of the system, such as in the solids feed systems, the gas seals, and the instrumentation.
- (b) Establishing the time required to achieve steady-state operation.

2. Develop a standard combustion test procedure allowing at least 2 hr of steady state operation in such test and including solids sampling to unambiguously determine the following: carbon burnout and combustion efficiency, sulphur capture, and solids elutriation.

3. Conduct combustion trials with Minto coal (high sulphur, from New Brunswick) and Havelock (New Brunswick) limestone to determine the effects of:

- (a) The Ca/S mole ratio
- (b) Bed temperatures of 750--950 °C
- (c) Bed height (expanded) of 0.3--1 m

- (d) Coal particle size
- (e) Limestone particle size
- (f) Excess air level

Because of the large number of variables, the initial tests were to be carried out using a two-level fractional factorial experimental design. Further tests were then to be done varying the dominant parameters and exploring any significant synergies with the aim of providing data to be used in predictive mathematical models of sulphur neutralization.

4. Conduct combustion trials with Devco coal and a limestone to be specified, to provide information to be used as a data base for the Summerside demonstration plant (Canadian Forces Base, Summerside, P.E.I.).

The shakedown runs were carried out first with sand beds and then with limestone. The greatest problem encountered was in achieving a durable and efficient pneumatic transport system for recycle of cyclone dust, an exceptionally abrasive material. After considerable trial and error, a very satisfactory system was developed.

The standard test procedure evolved out of the shakedown trials and the initial Minto trials.

The Minto trials were conducted as specified with one modification: only a single bed depth was used. This was partly because it was found necessary to first explore an important variable not mentioned in the above list, namely the recycle level. At the end, there was not time left for variation of the bed depth. In a certain sense, this is fortunate. The work that was done has, after close analysis, demonstrated how runs with varying bed depth should be designed in order to be productive.

For significantly deeper beds, more freeboard height and longer run times are required. Inclusion of such runs is planned for future work.

In the event, it was decided to also use the Havelock limestone in the Devco trials.

The planned and actual operating conditions of the Minto and Devco trials are set out in greater detail in Sect. 1.3.

1.2.3 Supporting studies specified
by the contract

In support of the pilot plant studies, the contract required:

- (a) The design, construction and commissioning of a bench-scale fluidized bed reactor for rapid preliminary investigations of fuels and limestones to supplement and help direct the pilot-scale work.
- (b) Development of a test procedure using thermogravimetric analysis (TGA) in conjunction with mathematical modelling to predict limestone performance. The pilot-scale studies would be utilized to indicate the limits of such predictive tools and the effects of any bed phenomena not taken into account in the operation of the TGA.

The bench-scale reactor, providing reaction chambers of 75 mm or 100 mm diameter, has been built and is being worked on in the thesis research of two graduate students, Mr. Kian McCaldon (M.Sc. program) and Mr. Jia-Qu Zhang (Ph.D. program). The study underway is on the burning of single large particles of coal, up to 40 mm dia. This information is greatly needed for mathematical modelling of the performance of FBC combustors burning, as in some of the present pilot scale trials, coal of large particle size.

The TGA work has been part of the thesis research of another graduate student, Mr. Michel Couturier (Ph.D. program), and is nearly complete. He has studied both the calcination of limestone and the reaction of the calcined particles with SO_2 and SO_3 . Other tools besides TGA have been employed, such as mercury porosimetry and electron microscopy.

Results of neither of these projects are included in the present report which is confined in scope to the Minto and Devco pilot plant trials. These results will appear in the student's theses which will also be made available as QFBC technical reports. The theses of Mr. McCaldon and Mr. Couturier are expected to be completed by December, 1983. Mr. Zhang, on the other hand, is at the start of his program, scheduled for completion in 1985.

1.2.4 Supporting studies outside the contract

A number of studies complementing those required by the contract have been underway, supported by an NSERC Strategic Energy Grant. These include:

- (a) A study of fluidization dynamics and heat transfer using a cold model having about $1/3$ the cross-sectional bed area of the pilot plant combustor, but provided with a similar distributor plate and in-bed tubes. This is part of the Ph.D. research of Mr. Fernando Preto.
- (b) A study of the mathematical modelling of FBC combustor performance, using the pilot plant results for test data. The model is a detailed mechanistic model. This is the second part of Mr. Preto's Ph.D. research.
- (c) A study of elutriation over very broad ranges of operation, using a cold model allowing continuous operation like that

in the pilot plant and served with spent bed material from the pilot plant. This has been the project of Mr. Bing-Song Xu, a Visiting Scholar from the People's Republic of China.

The results of Mr. Preto's work will appear in his thesis which is expected to be finished by April, 1984.

Mr. Xu has also involved himself in the handling of elutriation in the pilot plant trials. Both his cold-modelling data and the pilot plant data are treated in a paper on elutriation in FBC which has been prepared for the First International Fluidized Bed Combustion and Applied Technology Symposium in the People's Republic of China (Beijing, August 22-26, 1983). The paper is also being issued as a QFBC technical report, QFBC.TR.83.2. For this reason, the coverage of elutriation in the present report is confined to examination of a few salient findings and representative results.

1.3 The plan and actual operating conditions of the Minto and Devco trials

It was initially attempted to operate with overbed feeding of coal and limestone in which these materials are dropped or thrown into the bed at the top. This is the mode adopted for the first Canadian demonstration FBC boiler, located at Canadian Forces Base, Summerside, P.E.I. Because volatiles release tends to occur high in the bed, this mode can give intense freeboard burning. In our system as initially installed, it frequently produced exhaust gas cooling requirements exceeding the capacity of the pre-cyclone heat exchange equipment, overheating the cyclone and forcing shutdown. The widely fluctuating character of the volatiles release and freeboard burning also gave great problems in getting meaningful gas analyses. The PDP 11/10 computer had become useless at the time and the new computer was still months away in delivery. Thus computer-executed time-averaging of the half dozen signals simultaneously received from gas analysis was unavailable. To expedite progress, it was

therefore decided to adopt underbed feeding (injecting coal and limestone near the bottom of the bed) in the remaining trials under the present contract. Our subsequent experience indicates that this was a fortunate decision. The trials with underbed feeding have provided a good data base to which overbed feeding can subsequently be added as an interesting perturbation. Further, the effective freeboard height of the combustor should be extended as much as possible for overbed feeding in order to better accommodate the freeboard burning. Some trials with overbed feeding are planned for future work.

As already noted, we began work with a fully specified starting plan based on a two-level fractional factorial experimental design. However, compelling practical considerations quickly intruded and forced substantial changes. The experimental plan, it was found, had to allow a considerable element of trial and error and be subject to continual revision:

- o to begin with, the lots obtained of Minto coal and Havelock limestone were crushed and screened, and it was found that the lots produced of some size fractions were smaller than desired while others were considerably larger. Since funding was insufficient to obtain and process more coal and stone, the experiments had to be adjusted to the stock on hand
- o it was found that runs had to be longer than first expected in order to ensure steady-state operation, thus each run tended to consume more coal and stone than anticipated, forcing even stricter parsimony and care in the use of some size fractions
- o the first runs with underbed feeding, after the shakedown trials and the frustrations with overbed feeding, were done with coal and limestone of large particle size (as called for by the experimental design) and came to grief because of clinkering. It immediately became of highest priority to

find how, if possible, to avoid clinkering

- o the cyclone separator was soon found to be operating almost too efficiently, so that when cyclone dust was totally recycled the inventory of dust in the system rose catastrophically. The burden on the recycle system would in time exceed its capacity, plugging the line. Provision was then made for bleeding off a fraction of the cycloned dust, as product and the recycle fraction thereafter became another major variable to be studied. This was to the good because recycle is a significant variable and well worth a systematic investigation
- o because runs had to be long in order to ensure entry into steady-state operation, it was decided, in order to raise productivity (runs per week) to carry out runs in sets of two or three without intervening shut-down and startup. The same in-bed tube array was, of course, then used for all runs in a set. It was also often attractive to change only one variable from one run to the next, e.g., just the recycle level or the limestone feedrate.

For these various reasons, the record of experiments performed in the Minto trials soon differed conspicuously from the initial plan.

The Devco trials were very strongly influenced by the coal size distribution left after crushing and screening. The best that could be done with his equipment by our contractor resulted in a large portion oversize and only a small useable lot. Because of the lower sulphur content relative to Minto, the runs also had to be longer (because of lower limestone feedrate), consuming more coal per run. For these reasons, and because time and funds were not available to remedy the situation, only four runs with Devco coal were conducted. Taken in isolation, this gave rather scant information about the performance of this coal.

Taken in conjunction with the results on the Minto trials, however, the information is enough to reveal the essential differences.

A summary of the actual operating conditions of all the productive trials with underbed feeding is given in Table 1.1. It will be noted that we experienced some lack of success in setting the excess air at desired levels. There was a problem about the exact carbon content of the coals which made uncertain the prediction of excess air at given coal and air feedrates. The problem, discussed in Sect. 16.3, is that professional analysts at different laboratories seem to be unable to achieve reasonably close agreement on the carbon content of coal. Results reported on the same sample of Minto coal, Table 16.1, range from 66.4 % carbon to 71.4 % (d.b.). We are still investigating this question. For the present report, we have settled on using an average value as being the most likely to minimize possible errors. In the matter of setting excess air levels, however, we placed heavy reliance on measured exhaust gas oxygen contents, and it is evident that sometimes--because the carbon burnup was greater or less than anticipated--we settled outside the likely practical range. This is not necessarily bad since it broadened the range investigated and thus presumably helped to clearly define any significant effects.

1.4 Scope of the report

The foregoing discussion has already indicated some limits on the scope of the present report. In summary and completion of those remarks, we note that:

1. The report is confined to presentation of the results of the Minto and Devco pilot-scale combustion trials.
2. Results on supporting studies will be presented in other reports, theses and papers to follow.

Table 1.1. Operating conditions of productive runs. Footnotes identify noteworthy operating problems and information gaps. Bed temperatures are estimates of the mean steady-state values, duly emphasizing the late portions of runs.

Run	n_t	Coal		Stone		Ca/S	Air		r	T_b , °C	v, m/s	τ_p , hr	τ_b , hr	COT, hr
		D_p , mm	$\dot{m}_f$, kg/hr	D_p , mm	$\dot{m}_s$, kg/hr		$\dot{m}_a$, kg/hr	ϵ_a , %						
		Minto		Havelock										
820902 ^a	12	19.3	33.2	0.78	11	1.41	300	0	1	820	1.6	--	--	3.8
821116	12	19.3	29.3	0.78	8.4	1.25	260	-6.5	1	850	1.59	6.7	11.6	30
821130 ^b	24	19.3	50.0	0.78	23.8	2.03	383	-19.4	1	810	2.29	5.7	11.0	7.6
821206 ^c	14	10.7	26.5	1.98	13.5	2.17	255	6.3	1	870	1.60	13	15	8.7
830106 ^d	12	10.7	27.9	1.98	12.5	2.06	274	3.8	0.80	845	1.69	6	7.4	6.7
830112	14	10.7	31.0	0.78	11.4	1.60	280	-0.1	0.963	820	1.68	5.1	7.3	7.8
830124	12	10.7	32.8	0.78	15.8	2.10	302	-2.8	0.905	820	1.86	4.3	6.0	25
830126	12	10.7	30.3	0.78	28.0	4.25	293	1.9	0.836	805	1.75	3.0	4.7	12.2
830127 ^d	12	10.7	30.3	0.78	18.5	2.66	297	3.3	0.875	835	1.79	4.1	6.0	5.2
830201	12	6.6	33.3	0.78	17.5	2.24	311	-1.5	0.919	825	1.86	4.1	6.0	17.7
830202 ^e	12	6.6	33.1	0.78	28.0	3.60	320	2.0	0.906	840	1.90	2.8	3.9	4.7
830215	14	6.6	41.0	0.78	21.6	2.24	446	14.7	0.60	860	2.77	2.8	5.4	10.8
830216	14	6.6	40.9	0.78	28.0	2.92	446	14.9	0.52	865	2.76	2.4	4.4	10.7
830217	14	6.6	39.8	0.78	27.6	2.95	443	20.5	0	840	2.70	2.3	4.6	8.0
830218	14	19.3	38.0	1.98	23.3	2.70	447	24.0	0	860	2.77	2.9	4.6	9.7
830224	14	19.3	38.0	1.72	30.7	3.56	439	21.8	0.57	855	2.68	2.6	3.8	8.5
830225	14	6.6	39.0	1.72	26.3	3.10	441	19.3	0.54	940	2.93	3.1	5.2	9.7
830226	14	6.6	39.4	1.72	41.4	4.47	438	17.2	0.61	920	2.87	2.3	4.0	7.5
830322	12	19.3	31.1	0.49	15.1	2.14	298	1.2	0	840	2.14	3.5	8.2	9.7
830324	12	19.3	34.4	0.49	14.6	1.87	329	0.8	0.70	855	2.06	3.9	8.3	12.5
830329	12	19.3	38.2	0.49	31.0	3.73	357	-1.4	0.59	855	2.25	2.2	5.3	10.9
830330	12	19.3	37.4	0.49	30.3	3.57	322	-9.2	0	850	2.02	2.4	6.1	9.8
		Devco		Havelock										
830309	12	6.6	29.1	1.72	14.4	3.00	308	21.1	0.857	860	1.90	6.1	8.7	13
830310 ^d	12	6.6	27.1	1.72	10.2	2.28	306	29.6	0.76	870	1.90	8.6	13.6	5
830315	14	6.6	37.7	1.72	21.9	3.52	442	34.4	0.70	865	2.73	3.1	5.0	12.3
830316	14	6.6	35.2	1.72	13.2	2.27	441	43.6	0.76	850	2.67	4.6	7.3	13.7

^a Early shutdown due to breakdown in recycle system. Useful information obtained on heat transfer, but on no other aspect.

^b Early shutdown due to clinkering. Gas analyses not begun.

^c Recycle continually increasing until clinker formed, forcing shutdown. Gas analyses not begun.

^d Early shutdown because clinker diagnosed, subsequently disproven.

^e Early shutdown because of problem in baghouse inlet ducting.

3. Because of the great volume of the results, a limit has been set on interpretation. Detailed mechanistic analysis and mathematical modelling will be taken up subsequently in publications on various specialized aspects of the operation.

4. Again, because of the volume of the present results and also the great volume of the literature pertaining to fluidized bed combustion, discussion and interpretation of the present work in relation to the findings of others is strongly curtailed but will be pursued more closely elsewhere in publications dealing with specialized aspects of the operation.

1.5 Significance

The Queen's pilot plant project is part of an effort, originating in the Energy Research Laboratories, CANMET, Energy, Mines and Resources Canada (EMR) to promote and support the development and application of fluidized bed combustion in Canada. The effort is spurred by the fact that Canada has great resources of candidate fuels for FBC, and at the same time a need to strictly limit the emissions of sulphur dioxide from combustion processes. Other projects in the EMR program include:

- o a pilot plant like the present one, located at the Energy Research Laboratories of EMR, described in a report by Hector, Desai, Friedrich and Anthony (1981)
- o a pilot plant for investigating corrosion in in-bed tubes, located at Point Tupper, N.S., operated by Nova Scotia Power Corp.
- o the first Canadian demonstration plant, consisting of two FBC boilers, located at Canadian Forces Base, Summerside, P.E.I. Two reports on this project are available (Taylor and Friedrich 1982; Razbin and Friedrich 1983).

Canada, through EMR, is a participant in the International Energy

Agency (IEA) Agreement for Cooperation in the Field of Atmospheric Fluidized Bed Combustion in Industrial or District Heating Boilers. The present project thus contributes, as well, to this international cooperative effort. Descriptions of all the projects in progress in the countries taking part in the IEA agreement may be found in the annual reports of the Executive Committee for the agreement.

The present endeavour also has significance beyond carrying out research required by EMR. In the engineers, technicians and graduate students involved, professionals are being trained for work in the technology. A Canadian centre of expertise and investigative resources complementing those at EMR has been established. Finally, for its part in supporting graduate students, bench scale studies, and theoretical work through a Strategic Energy Grant, and providing instruments such as the TGA apparatus through equipment grants, useful work is being done in an area of research, namely energy, that has been designated as being of strategic national importance by the Natural Sciences and Engineering Research Council of Canada.

2. THE QUEEN'S AFBC PILOT PLANT FACILITY

2.1 Introduction

The AFBC pilot plant is described in a recent technical report (Becker & Code 1982) hereinafter referred to as QFBC.TR.82.1. The presentation here is therefore limited to a brief review of salient features of the system. Several minor modifications and new components introduced after issuing the said report are also described, and some planned future modifications and additions are noted. The whole pilot plant in its present form is shown schematically in Fig. 2.1.

2.2 The combustor

The heart of the system is the fluidized bed combustor, Fig. 2.2, which consists of nine modules. The chamber cross-section is nearly square, 381 mm by 406 mm, 0.155 m^2 cross-sectional area, in all modules except the top which is transitional. From bottom to top, the modules are:

1. The windbox, a new, improved unit installed after QFBC.TR.82.1 was issued. It is light in weight and attaches to the assembly above (to module 3) by just four bolts, facilitating rapid removal and reattachment. The internal chamber height is 254 mm and the volume is 0.039 m^3 . Air is fed through a 75 mm dia port in the middle of one side and leaves through the roof which is constructed to be the distributor plate serving air to the bed of particles overhead. This plate is fitted with 100 distributor heads uniformly spaced on a rectangular pitch 37.5 mm by 37.8 mm, Fig. 2.3. As before (QFBC.TR.82.1), the distributor heads are 12.7 mm dia stainless steel cap screws drilled with four horizontal discharge holes just under the cap and a feed hole up the shaft. It should be noted that the windbox is regularly dropped after each run or series of connected runs, for a number of purposes: to dump the bed for weighing (determination of holdup) and examination, to remove accumulated

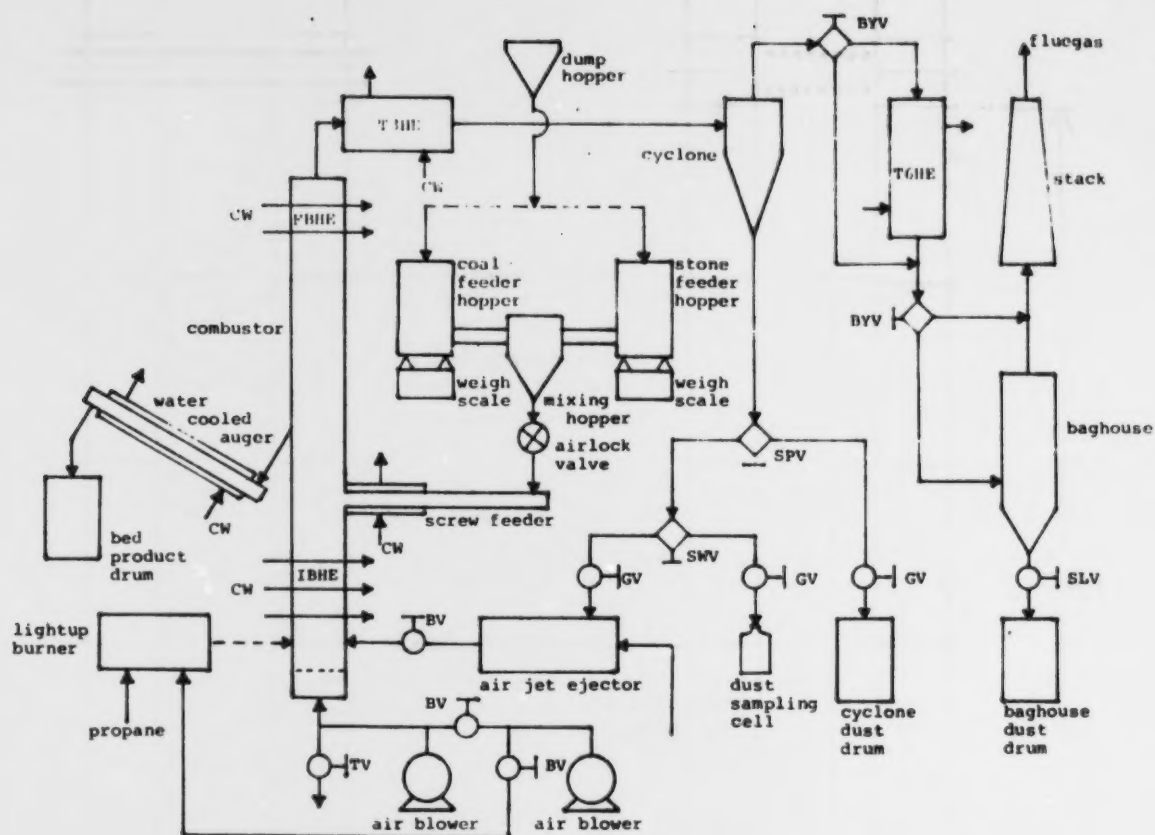


Fig. 2.1. Flowsheet of the AFBC pilot plant. Acronyms:

BV	ball valve
BYV	by-pass valve
CW	cooling water
FBHE	freeboard heat exchanger
GV	gate valve
HP	high pressure
IBHE	in-bed heat exchanger
SLV	slide lock valve
SPV	splitter valve
SWV	switch valve
TV	throttling valve
T3HE	heat exchanger cooling transfer stream T3
T6HE	heat exchanger cooling transfer stream T6

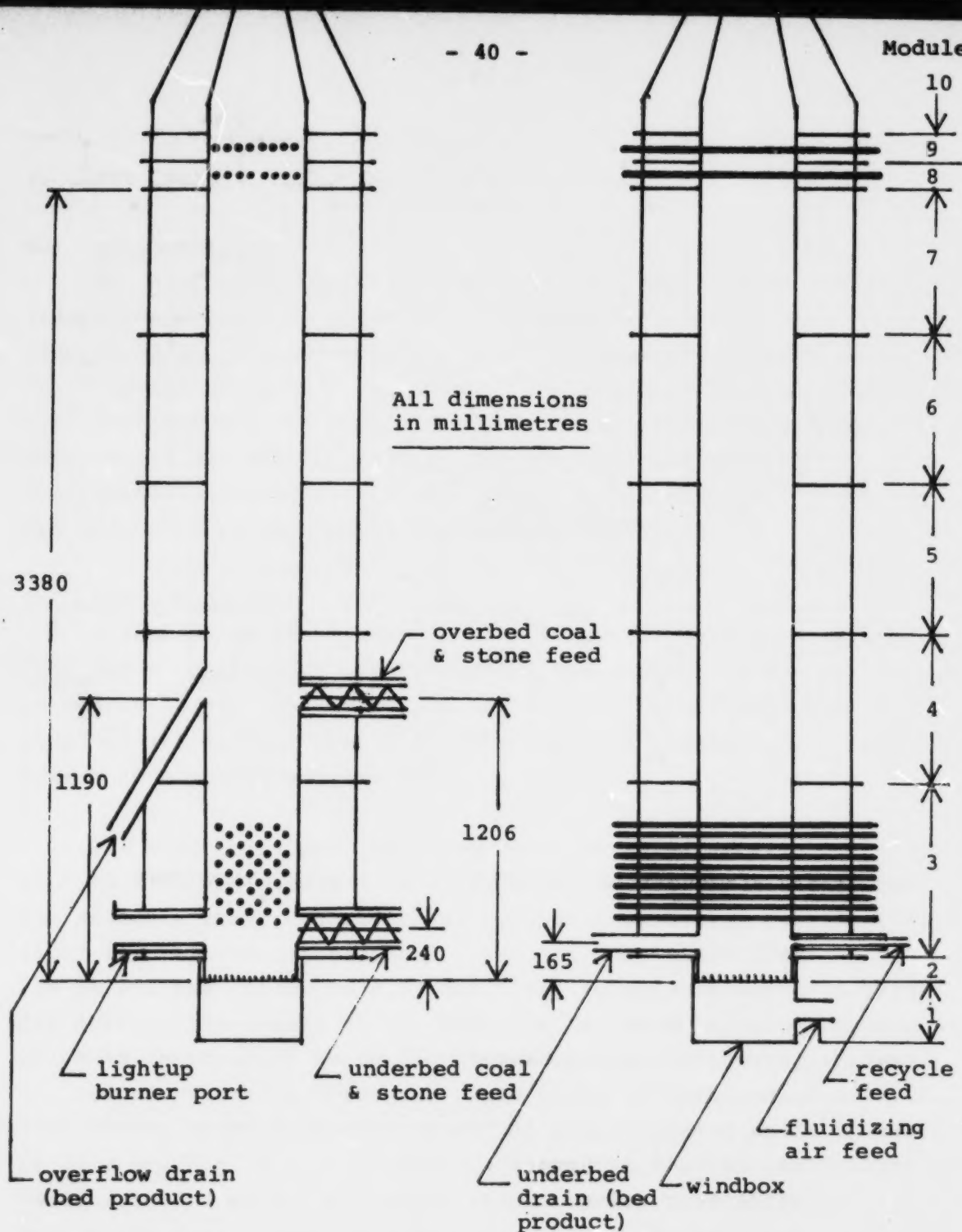


Fig. 2.2. Principal dimensions and features of the combustor

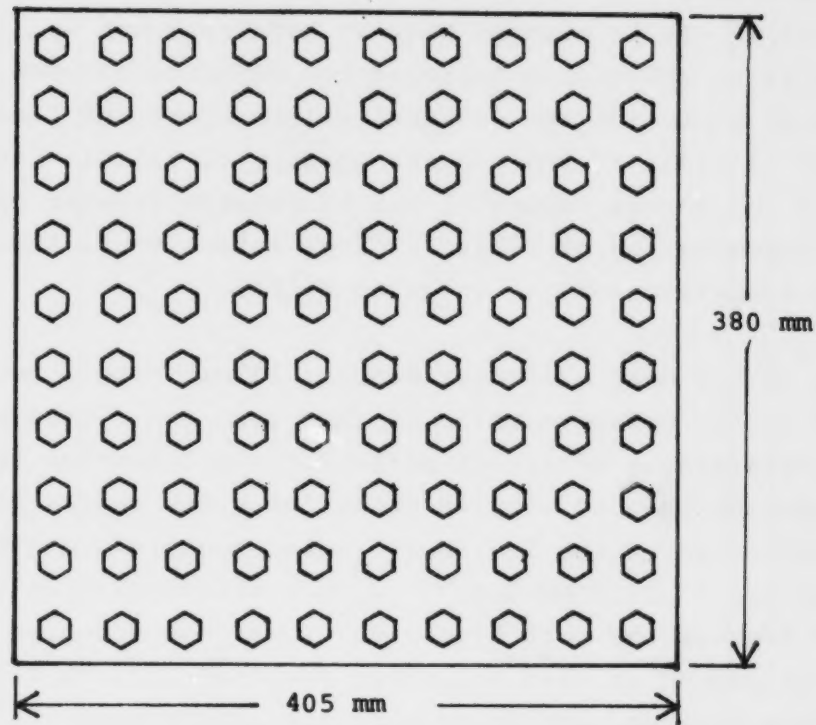


Fig. 2.3. Pattern of the distributor heads on the distributor plate

"rocks" (unfluidizably large stone particles), and to allow examination of the in-bed heat-exchange tubes and of the refractory walls and linings.

2. A waterwall downward extension, 102 mm high, added after QFBC.TR.82.1. It is clamped between modules 1 and 3. Its function is to reasonably increase the distance between the distributor plate and the underbed solids (coal and limestone) feedport, in order to improve the quality of fluidization and mixing at the solids feedpoint and to provide greater capacity for the accumulation of rocks (if any) before these interfere with the dispersion of the entering solids.

3. The in-bed heat exchanger module, 737 mm high. Rectangular openings in two opposite walls of this unit are fitted with panels containing a matrix of ports for the insertion of cooling tubes spanning the bed when in operation. The panels are refractory-lined on the inside and water-jacketed on the outside. The tubes are of 21.3 mm O.D. (1/2-inch standard) steel pipe. Figure 2.4 shows the port geometry. Ports not occupied by tubes are either capped or used for other purposes such as sampling, temperature probing, and feeding solids. This module also contains the ports, Fig. 2.2, for underbed feeding of coal and limestone, connection of the lightup burner, pneumatic injection of recycling cycloned dust, and (not used in the trials described herein) underbed discharge of bed solids.

4--7. Four internally identical modules, each 635 mm high. The lowest has, Fig. 2.2, on opposite sides, two solids transport ports: a bed overflow port allowing overflow of bed solids into a 77.9 mm I.D. (3 inch std.) drain pipe set at 30° to the vertical, and an overbed feed port for coal and limestone. These modules, and module 3 as well, are all fitted with numerous ports for insertion of sampling probes, thermocouples, and pressure taps (details given in QFBC.TR.82.1).

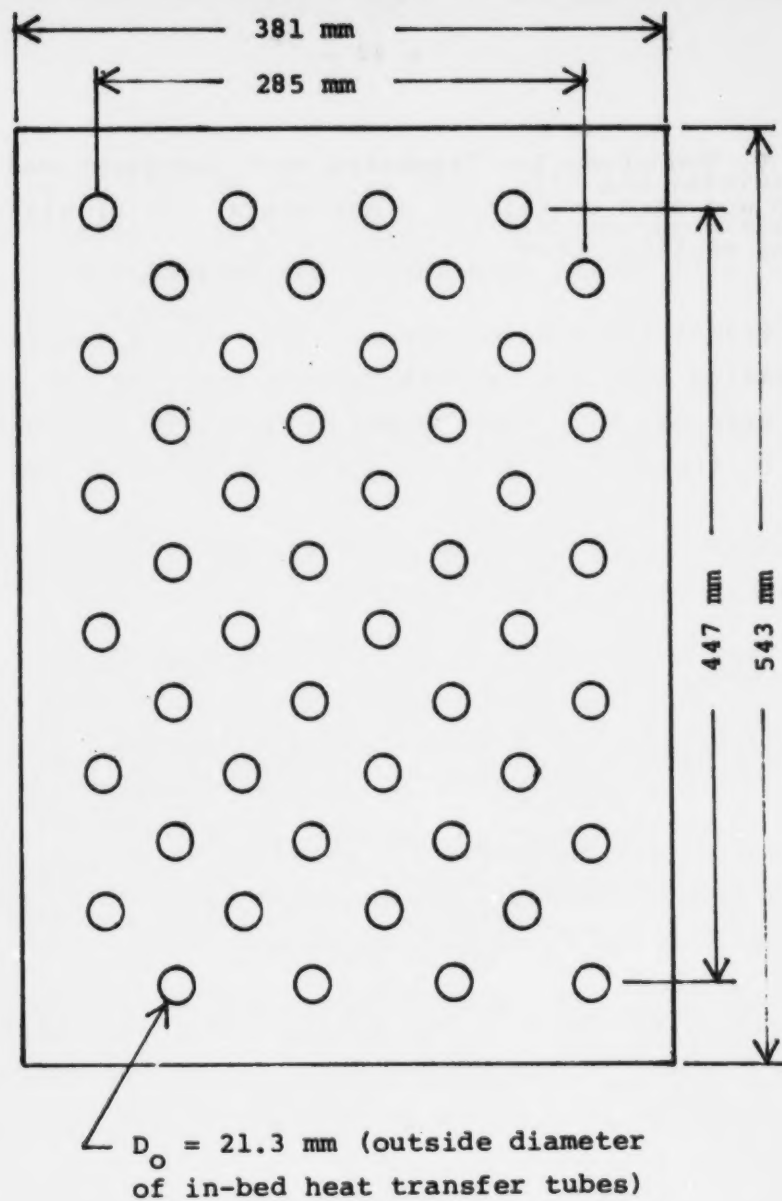


Fig. 2.4. Distribution of ports for in-bed heat exchange tubes

8. & 9. Two crossflow freeboard heat exchanger modules, each 120 mm high and each containing a horizontal row of eight tubes carrying cooling water.

10. A transition module, 610 mm high, with a contraction at the end, leading into the exhaust ducting carrying the exhaust gases to the external heat exchangers, cyclone and baghouse. This module is fitted with ports for exhaust-gas sampling.

Modules 3--10 are made up of 231 mm thick refractory brick wall contained in a steel shell.

In summary:

- o the combustor has provision for overbed and underbed feeding.
- o bed solids may be removed by overflow or by underbed extraction. These are the means of controlling the solids charge ("holdup"), or the bed height.
- o heat is extracted from the bed via the in-bed heat exchanger tubes (in module 3), the waterwall extension (module 2), and the water-jacketed panels accommodating the in-bed tubes in module 3.
- o recycling cycloned dust may be injected into the bed through the port in module 3 indicated in Fig. 2.2 or through any of the unused ports for in-bed heat exchanger tubes.
- o heat is extracted from the gases at the end of the freeboard by the freeboard heat exchanger.
- o the height of the combustion chamber from the distributor plate at its bottom to the first row of freeboard heat exchanger tubes at the top with the waterwall extension (module 2) in place, is 3.34 m.

- o the bottom rim of the bed overflow port, with the waterwall extension in place, is 1.19 m above the distributor plate. This, then, is the nominal bed height when holdup is controlled by overflow.

2.3 The exhaust gas train

The exhaust gases pass through the following units on their way to the stack:

1. A short duct fitted with an explosion vent
2. A shell-and-tube heat exchanger with the gases on the tube-side and cooling water in the shell-side, labelled T3HE (heat exchanger cooling transfer stream T3, Fig. N.1). This drops the temperature below appreciable chemical reaction levels and to a level suitable for the cyclone separator, below 425 °C.
3. The cyclone dust separator.
4. A cross-flow heat exchanger with the gases in the shell, water in the tubes, labelled T6HE (heat exchanger cooling transfer stream T6, Fig. N.1). This provides control of the gas temperature entering the baghouse and going to the stack.
5. The baghouse.

The explosion vent and the gas quenching heat exchanger are new, installed after QFBC.TR.82.1. The explosion vent has a disk or lid firmly pressed over a 152 mm dia hole by weights suspended on wires. The weights and the wires are suitably sized so that in the event of an explosion the wires snap, the lid swivels up on a hinge, and the weights are transferred to restraining chains. The heat exchanger T3HE incorporates a bundle of nineteen 48.3 mm O.D. (1-1/2 inch std. pipe), 1.52 m long, tubes.

A bypass valve permits partial by-passing of heat exchanger T6HE, so the gases going to the baghouse will not be overcooled. Another bypass valve permits bypassing of the baghouse until the gas temperature for entering the baghouse has been suitably adjusted. Details are given in QFBC.TR.82.1.

2.4 The coal and limestone feed systems

Coal and limestone are fed from separate feeder-hoppers mounted on electronic weigh scales. The two streams fall into a tee connection between the feeders and continue together through a rotary air-lock valve ("star feeder") and on into a screw feeder which expells them into the bed at the inside wall of the combustor. Details are given in QFBC.TR.82.1.

2.5 The cyclone dust recycle system

The dust stream leaving the cyclone is split in two by an adjustable splitter valve. One part, of mass flux $\dot{m}_r$, is recycled. The other, of mass flux $\dot{m}_c$, is drawn off as product. The recycle fraction $r = \dot{m}_r / \dot{m}_d$ is continuously variable, by means of the valve, from zero to unity, where $\dot{m}_d = \dot{m}_r + \dot{m}_c$ is the total mass flux of dust leaving the cyclone. This adjustable splitter valve was added after issuing QFBC.TR.82.1. The previous valving was "on-off"; either none or all of the dust was recycled.

The recycle dust stream flows, by gravity and suction, into an air-jet ejector which acts as the pump for pneumatic reinjection into the bed. A switch valve just before the ejector allows diversion of the recycle stream to a sampling cup, for material sampling and for determination of flow rate by collecting the stream for a measured time interval.

The terminal parts of the recycle line are of 1/2-inch standard (15.8 mm O.D.) stainless steel pipe. These parts are short and straight. The long curved section connecting them is of 19.6 mm I.D. rubber sand-blasting hose.

2.6 Air supply and flows

Fluidizing air, fed to the windbox and entering the bed through the distributor plate, is supplied by a Roots blower. Dust recycle air is supplied from the laboratory high-pressure air line, as is a small flow of air injected into the coal/limestone feed line to offset gas leakage through the rotary air lock valve and to block escape of bed gases through that valve. The recycle air flux in the work described herein was normally about 18 kg/hr. The spillage of purge air into the combustor from under the air-lock valve was 2 kg/hr. The fluidizing air flux was 250--450 kg/hr and made up 90--95 % of the air supplied.

2.7 Instrumentation and analytical systems

The gas and solids sampling probes, the gas and solids analysis systems, the temperature and pressure measurement systems, and the data acquisition and processing system (based on a desk-top minicomputer) are all described in QFBC.TR.82.1. In brief, infrared analyzers were used for on-line measurement of CO, CO₂ and SO₂ concentrations, a paramagnetic analyzer for O₂, a chemiluminescence analyzer for NO and NO_x, and a flame-ionization analyzer for unburnt hydrocarbon ("HC", indicated as ppm of CH₄).

Thermocouples were used for temperature measurement, doubled at critical points in the air and water circuits by dial thermometers.

Proximate analysis and sulphur analysis of solids were done in our laboratories. Analyses for carbon, hydrogen and nitrogen were done in commercial laboratories.

3. OPERATING PROCEDURES

3.1 Cold startup

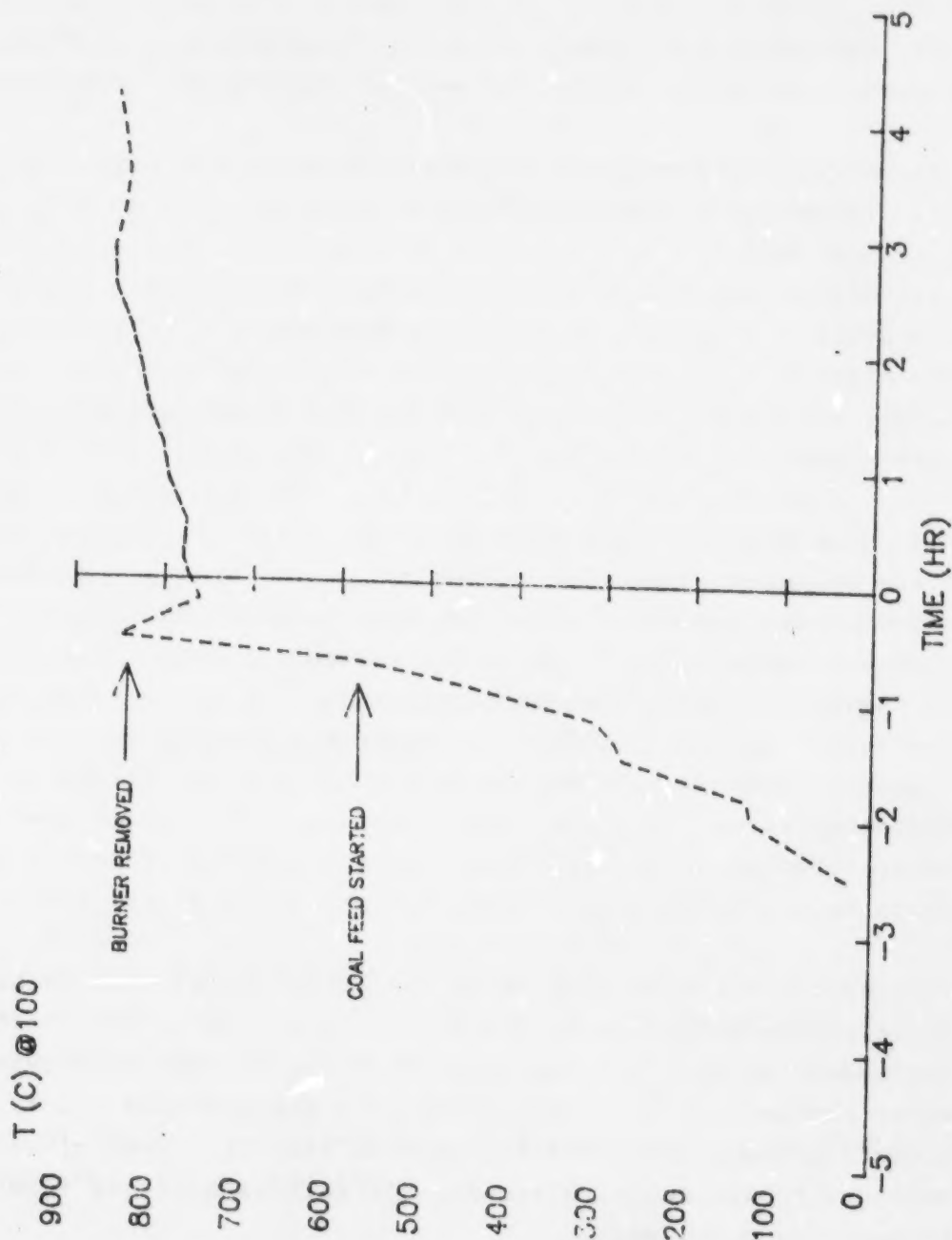
The normal procedure for startup from cold conditions was as follows:

1. Attach the lightup burner to the combustor.
2. Charge the combustor with bed material from a previous run (or, lacking that, with limestone), in quantity 80--100 % of the expected steady-state holdup. To minimize the time required to reach steady state, the material should be as similar as possible to that expected from the current run.
3. Fully close the pre-baghouse heat exchanger (T6HE) bypass valve, Fig. 2.1, directing the whole exhaust gas flow through this exchanger.
4. Fully open the baghouse bypass valve, Fig. 2.1, directing the whole exhaust gas flow around the baghouse.
5. Establish cooling water flows to all heat exchangers and cooling jackets.
6. Establish a fluidizing airflow sufficient to give good agitation, typically a fluidizing velocity under cold conditions around 0.4 m/s (corresponding to 1.5 m/s at 850 °C).
7. Establish normal operating flows of recycle air and solids feedline (air-lock valve) purge air.
8. Start the lightup burner at low fire. Increase the firing rate at 150 °C and 300 °C bed temperature.
9. Commence feeding coal and limestone at 600 °C bed temperature.

Fig. 3.1. Typical record of bed temperature in startup from cold conditions

RUN 830324 START-UP TEMPERATURE

(TIME=0 AT INSTRUMENTATION ON)



10. When it reaches 850 °C, slump the bed (turn off the fluidizing and recycle air flows), quickly disconnect the lightup burner, and cover its port in the combustor with a blind flange.

11. Restart the airflows, with the fluidizing air at the expected operating level. Recommence feeding coal and limestone.

12. Attach the bed product cooler (water-cooled auger) and, at its discharge, a product collection barrel.

13. Adjust the air, coal and limestone feedrates to produce the desired steady-state operating conditions. Most commonly, in the present work, the operating bed temperature, excess air level, and Ca/S ratio were fixed and the flows were adjusted to meet these.

14. When the bed depth reaches its expected (or planned) steady-state value, as indicated by the bed pressure drop, open the valve at the operating combustor bed-solids discharge port and commence removing bed material to maintain constant bed depth. In the present work, bed depth was regulated by overflow, using the overflow port located 1.2 m above the distributor. Note that limestone feeding in steps 7--12 did not always follow the indicated normal pattern; the limestone feedrate was generally manipulated as necessary to produce the expected steady-state bed depth before being adjusted to give the desired Ca/S ratio.

15. Adjust the bypass valve on the pre-baghouse heat exchanger (partly bypassing some of the exhaust gas flow around the heat exchanger) so that the temperature of the recombined streams after the exchanger is appropriate for the baghouse (normal operating temperature 150 °C, maximum 200 °C). Then close the baghouse bypass valve, directing the full exhaust gas flow through the baghouse.

16. In startup and in subsequent operation, adjust all cooling water flows so exit temperatures are no higher than 60 °C. This rule ensured avoidance of boiling. Finally, adjust the cooling water flowrates through the bed-cooling heat exchange devices (in-bed tubes, combustor tube-wall cooling jackets) so as to allow reasonably accurate readings of both water flowrates and temperature rises.

Figure 3.1 shows a typical record of bed temperature for the startup period from a cold start. The time required to reach operating temperature was typically 2-1/2 to 3 hr. Higher rates of temperature rise are attainable but dangerous to the refractory walls of the combustor.

3.2 Hot start of a run

Whenever possible, several runs were conducted in series without stoppage between runs. This saved the time otherwise required for startup from cold conditions and more than doubled production (runs made per week). The input variables changed from one run to the next were then typically limestone particle size, Ca/S ratio (through limestone feedrate), coal particle size, bed temperature, and/or cyclone dust recycle ratio. The coal burning rate could not be changed significantly, the bed cooling surface (primarily the in-bed heat exchange tubes) being fixed.

3.3 Procedure during runs, after startup

Full data collection generally began with the start of bed-solids removal. Thereafter the following variables were recorded at suitably close intervals:

- o weights of the coal and limestone feeder hoppers, to determine the feedrates of coal and stone
- o cooling water temperatures and flowrates

- o air temperatures and data for determining air flowrates (manometer readings indicating pressure drop across orifices, etc.)
- o solid effluent production (bed solids, cyclone dust, baghouse dust). Bed solids were collected in a barrel for a period of 1-1/2 to 3 hr (depending on the production rate and the planned duration of the run). The barrel was then replaced and the mass of bed material collected was determined by weighing. The same procedure was followed with the cyclone product. The baghouse product, however, accumulated in the baghouse hopper over the entire duration of a run and was dumped and weighed only at the end. The production rate of baghouse dust was usually low (1--3 kg/hr) and the baghouse dust collection hopper was thus too crude a device to allow meaningful determination of that rate by dumping at relatively short intervals.
- o the mass cyclone dust recycle rate. Determined by diverting the recycle flow into a sampling cell for a short time. Sampling times ranged from 15 s to 180 s.
- o gas composition at combustor exit (O_2 , CO, CO_2 , SO_2 , NO, NO_x , unburnt hydrocarbon)
- o temperatures in bed and freeboard and at other critical points in the system as indicated by fixed thermocouple probes
- o static pressures in the windbox, at several levels in the bed, and at other significant points in the system.

When it was judged that steady state was near or effectively there, gas composition was determined at various levels in the freeboard with a gas sampling probe, and particulate samples were taken with a particulates sampling probe. Samples of bed material were taken from several points in the bed.

Just before the end of a run, an attempt was sometimes made to determine the minimum fluidizing velocity. The recycle air flow was shut off. The fluidizing air flow was then reduced by steps and the static pressure drop across the bed was recorded as a function of flowrate. This procedure had to be carried out as rapidly as possible in order to avoid excessive change in bed temperature.

3.4 Criteria for the approach of steady state conditions

To ensure that runs were carried on sufficiently long so that the steady state was approached closely enough to yield results representing that state with negligible error, emphasis was placed on the length of operating time relative to the residence time of solids in the bed. An obvious estimate of such a residence time is

$$\tau_p = m_b / \dot{m}_p, \quad (3.1)$$

$$\dot{m}_p = \dot{m}_b + \dot{m}_c + \dot{m}_h,$$

where m_b is the mass of solids in the bed (the mass holdup) and $\dot{m}_p$ is the total mass efflux of solid products, the sum of the bed product flux $\dot{m}_b$ (bed material withdrawn by overflow or mechanical extraction), the cyclone product flux $\dot{m}_c$, and the baghouse product flux $\dot{m}_h$. This estimate is, however, rather naive. A worst-case estimate would have it that bed product, cyclone product, and filter product all have different residence times, but those of the latter two are very small and their contributions to the holdup are correspondingly negligible. Hence emphasis should be placed on the residence time of bed product,

$$\tau_b = m_b / \dot{m}_b. \quad (3.2)$$

One of the useful features of gas-fluidized beds is the vigorous agitation and mixing of the bed produced by

fluidization. For the present argument, it is reasonable to suppose that the mixing is virtually perfect. The residence time distribution of bed product is then

$$F(\tau) = \exp(-\tau/\tau_b), \quad (3.3)$$

where $F(\tau)$ is the probability that a particle will stay in the bed for a time τ after entering it (or, for a collection of particles leaving the bed, $F(\tau)$ is the fraction that were in the bed for a time less than τ). The consequence of this is that if we consider the particles that are in the bed at a given time t , a mass fraction $1/e = 0.368$ of these will still be there at time $t + \tau_b$. After two residence times τ_b , the fraction remaining is $1/e^2 = 0.135$, and after three it is $1/e^3 = 0.050$. One may thus conclude that if t is the operating time from the instant when the approach to steady state is begun, then at $t + 2\tau_b$ the system must be very close to steady state, and at $t + 3\tau_b$ it should be fully arrived by any reasonable definition. This is a valid conclusion for our system because the establishment of a full population of particles with steady-state characteristics, which effectively happens when the initial population has been flushed out, is the slowest step in the approach to steady state. Other processes, the slowest of which is the heating of the refractory combustor walls to the point where they take negligible heat from the bed, are all considerably faster. In this argument, the starting time for operation towards the steady state may be practically regarded as the time at which removal of bed material is begun and bed turnover thus initiated. At this time the bed temperature in our work was generally near the desired level, and the coal and limestone feedrates were at or near their final set points. The act of starting continuous removal of material from the bed to maintain a constant holdup marked the initiation of fully continuous operation. We shall hereinafter call the operating time measured from this instant the COT time (Continuous-mode Operating Time).

Our results (see Sect. 5.4) show that the above criterion for attainment of steady state conditions was always sufficient, and often more than enough. For example, when a run was begun hot, with no significant interruption between it and the preceding run, and the change of operating conditions was a comparatively mild one such as a change of coal particle size, then the system effectively equilibrated after a COT time lapse as short as one residence time τ_b . However, it was also clear that the rule of "steady state at $2\tau_b$ -- $3\tau_b$ " should normally be adhered to. In-run indications of steady state from temperatures and gas analyses can be seriously misleading. As indicated above, it is the particle characteristics that take longest to equilibrate, and since the solids chemical analyses could not be done during a run, only after, the final and decisive judgment on the effective arrival of steady state could only be made retrospectively.

The question of reaching the steady state is of great importance. Practically, it is steady-state data that are most wanted and needed. Transient-state data are also desired, of course, in regard to questions of control, stability, load-following, fuel-changing, etc., but then the experiments should be designed to answer questions about the particular types of transients that are of interest.

3.5 Post-shutdown measurements

When the system had been shut down and cooled, the windbox was normally dropped and the bed material was collected and weighed. A sieve analysis was also done to see if the particle size distribution differed from that of the material collected as bed product during the run. Significant enrichment in large particles indicates stratification during operation, with accumulation at the bottom of the bed of large unfluidizable or poorly fluidized particles.

3.6 Manpower and scheduling considerations

The frequent need for long runs to ensure close approach to steady state, and the efficient practice of conducting several runs in series without stoppages between, meant that manpower had to be deployed in shifts. The minimum shift size was three workers. The size used was normally three or four. Because the normally available manpower was sufficient only for two shifts, a twelve-hour shift length was usually used. The operating time per week was then between 24 and 72 hours, depending on the nature of the runs involved. When feeding coal of large particle diameter, for example, accumulation of "rocks" (unfluidizable large stone particles) in the bed usually forced relatively early shut-down to allow removal of this material.

4. MATERIALS: COAL AND LIMESTONE

4.1 Coal

4.1.1 General characteristics

An evaluation of Nova Scotia and New Brunswick coals has been made by Tibbetts and Montgomery (1976). Both Minto coal (central New Brunswick) and Devco (Cape Breton, Nova Scotia) are ranked as high volatile A bituminous (ASTM classification). They are fairly strongly swelling, the range of the free swelling index being 5--5 1/2 for Minto and 5--8 for Devco. Ash contents are in the range 15--25 % for both. The sulphur content of Minto coal is typically 7--8 %, but that of Devco, 1--5 %, varies considerably between different mines.

4.1.2 Origin and handling of the present lots of coal

The Minto coal was purchased from a mine operator (New Brunswick Coal Ltd., Minto, N.B.) as run-of-mine coal, containing lumps up to 20 cm dia, hauled to Kingston, and crushed (in a roll mill) and screened by a Kingston quarry operator. The Devco coal, which came run-of-mine with lumps up to 7 cm dia, was similarly purchased (from Cape Breton Development Corporation, Sydney Mines, Sydney, N.S.) and handled. The Devco coal was from the Prince Mine of C.B.D.C.

4.1.3 Handling of chemical analyses

Proximate analyses and sulphur analyses were performed in our laboratory. Carbon, hydrogen, nitrogen (CHN) analyses were routinely performed by a commercial analyst (referred to hereinafter as analyst A), as were analyses for ash components. Coal was stored and moved in barrels. Because of the time and expense involved in sample preparation and analysis, it was impractical to sample every barrel. Instead, a number of barrel lots of coal were carefully sampled and analyzed and the average analyses computed from these results were used in all our

material balances and other computations. The coal samples were also sent to a number of other analysts besides analyst A for CHN analysis, and all these analyses were critically considered in arriving at the average analysis. A comparison of the results found by the different analysts is given in Sect. 16.3.

4.1.4 Proximate analyses

The moisture content (dry basis) of the Minto coal was 0.75 ± 0.05 %, and that of the Devco was 2.4 ± 0.1 %. The other results of the proximate analyses are summarized in Table 4.1 in which comparisons are also shown with representative literature data. The ratio of fixed combustible matter (often referred to as "fixed carbon") to total combustible matter, FCM/CM, is evidently a virtually constant characteristic for coal from a given field:

- o Minto, present data: 0.594 ± 0.007
- o Minto, Tibbetts and Montgomery (1976), ten samples from four mine sites: 0.597 ± 0.015
- o Devco (Prince Mine), present data: 0.581 ± 0.002
- o Devco (Prince Mine), data furnished by Cape Breton Development Corporation: 0.587
- o Devco, Tibbetts and Montgomery (1976): 0.591 ± 0.005 for Sydney, Point Aconi and Lingan mines, and 0.630 ± 0.008 for Glace Bay mines.

It appears, too, that the different analysts are in good agreement on the relative proportions of volatile combustible matter and fixed combustible matter.

4.1.5 Ultimate analyses

Ultimate analyses are summarized in Table 4.2, again with

Table 4.1. Proximate analyses of Minto and Devco coals, dry basis

Coal	Minto ^a	Minto ^b	Devco ^c	Devco ^d
Reporter	Queen's	TM	Queen's	CBDC
VCM, %	33.7 ± 0.2	32.5	33.6	34.7
FCM, %	49.3 ± 0.9	48.8	46.5	49.3
Ash, %	17.4 ± 1.4	18.8	20.0	16.1
FCM/CM	0.594 ± 0.007	0.600	0.581 ± 0.002	0.589

^a The present lot of Minto coal, average of nine analyses
(Three samples each for the three particle sizes)

^b Tibbetts and Montgomery (1976), typical run-of-mine sample
(ERL no. 2073-76)

^c The present lot of Devco coal (Prince Mine), average of
three samples

^d Analysis of Devco coal (Prince Mine) furnished to us by
Cape Breton Development Corporation, based on product over
one week in 1981

Table 4.2. Ultimate analysis of the coal dry matter. Oxygen, O, is estimated by difference.

Coal	Minto ^a	Minto ^b	Devco ^a	Devco ^c
Reporter	Composite	TM	Composite	CBDC
C, %	68.3	66.3	64.7	67.8
H, %	4.1	4.3	4.1	5.1
N, %	0.8	0.9	1.1	1.26
S, %	7.4	8.0	5.4	4.8
Ash, %	17.4	18.7	20.0	16.0
O, %	2.0	1.8	4.7	5.0

^a Best estimate, based on analysis of our coals by several analysts. These are the analyses used in our element balances.

^b Tibbetts and Montgomery (1976), typical run-of-mine sample (ERL no. 2073-76).

^c Analyses for Devco coal (Prince Mine) furnished to us by Cape Breton Development Corporation, based on product over one week in 1981.

comparisons with published results. The oxygen contents, obtained by difference, must be regarded with some reservation. The ash, for example, presumably contains any sulphate sulphur originally present in the coal. The sulphur analysis, however, gives the total sulphur. Thus the sulphate sulphur is counted twice. The analysis of Devco coal (Prince Mine) that was given us by Cape Breton Development Corporation indicates that about 2 % of the total sulphur in it is sulphate sulphur, so our Devco coal is 0.11 % sulphate sulphur and 5.3 % non-sulphate sulphur. This lowers the estimate of oxygen from 4.7 % to 4.6 %. The error in this is small, but more seriously, all errors in the ultimate analysis are combined in the estimate of the oxygen. Incidentally, another consequence of part of the coal's sulphur being in sulphate form is that this sulphur presumably remains fixed in the ash in fluidized bed combustion. This makes it appear that more sulphur is captured by the limestone than is actually the case. However, for the present Devco coal and presumably for the Minto, the proportion of sulphate sulphur in the coal is not high enough to be important.

4.1.6 Ash analyses

Ash analyses obtained by our commercial analyst A and published results of others are shown in Table 4.3. The levels of Mg, Na and K found by analyst A are very low and may be in serious error. The amount of iron found by him also seems low. We are not satisfied with his findings for Mg, Na and K, but have had to use them in the element balances anyway (Sect. 12) because he also analyzed the product streams. This analytical problem will be resolved in future work.

4.1.7 Ash fusibility

The ash of Minto coal, Table 4.4, softens at relatively low temperatures. This suggests that clinkering might be a problem under adverse conditions in fluidized bed combustion, and it indeed was found to be so. The temperatures characterizing fusibility are all around 200 °C higher for Devco coal, high

Table 4.3. Ash analysis of Minto and Devco coals.

Coal	Minto ^a	Minto ^b	Devco ^a	Devco ^c
Reporter	Analyst A	TM	Analyst A	CBDC
SiO ₂ , %	41.6	28.3	41.6	39.1
Al ₂ O ₃ , %	10.3	11.1	23.8	22.3
Fe ₂ O ₃ , %	37.4	43.2	23.7	31.2
CaO, %	3.9	3.8	1.00	1.39
MgO, %	0.05	0.8	0.05	1.14
Na ₂ O, %	0.2	2.8	0.15	0.53
K ₂ O, %	0.1	0.9	0.05	1.76
TiO ₂ , %	--	0.7	--	0.30
P ₂ O ₅ , %	--	2.3	--	0.30
SO ₃ , %	--	3.5	--	1.42
Total det.	93.6	97.4	90.4	99.43

^a One sample

^b Tibbetts & Montgomery (1976), typical run-of-mine sample (ERL no. 2073-76)

^c Analysis for Devco coal (Prince Mine) furnished to us by Cape Breton Development Corporation, based on product over one week in 1981

Table 4.4. Ash fusion characteristics for Minto and Devco coals.
Note:

IT = initial temperature
ST = spherical softening temperature
HT = hemispherical softening temperature
FT = fluid temperature

Coal	Minto ^a	Devco ^b	Devco ^b
Reporter	TM	CBDC	CBDC
Conditions	Oxidizing	Oxidizing	Reducing
IT, °C	1066	1259	1091
ST, °C	1099	1319	1141
HT, °C	1138	1368	1186
FT, °C	1215	1406	1310

^a Tibbetts and Montgomery (1976), typical run-of-mine sample (ERL no. 2073-76)

^b Data for Devco coal (Prince Mine) furnished to us by Cape Breton Development Corporation, based on product over one week in 1981

enough so that clinkering should not normally be a danger.

4.1.8 Calorific value

Calorific value was determined in our laboratory by means of a Parr combustion calorimeter, now unstylish and relatively very tedious to run, but still serviceable. The value found, Table 4.5, for the calorific value of Minto coal on an ash-free, moisture-free basis (thus the calorific value of the combustible matter) was 34.6 ± 0.4 MJ/kg. This agrees closely with the value of 34.9 ± 0.3 MJ/kg given by the results of Tibbetts and Montgomery (1976) for ten samples from the Minto field. For the Devco coal we obtained 31.5 ± 0.7 MJ/kg, compared to 33.54 MJ/kg indicated by the data given us by the supplier (Cape Breton Development Corporation). The results of Tibbetts and Montgomery (1976) for 24 samples of Devco coals from various mines give 35.6 ± 0.5 MJ/kg. The calorific value of pure carbon is 32.77 MJ/kg, and a value for coal less than this is unlikely. We observed that the Devco coal tended to be unreactive in the determination and a quantity of benzoic acid had to be added to samples to promote burning. We evidently failed to solve the problem completely. Based on the consistent results of Tibbetts and Montgomery, the calorific value (on ash-free, moisture-free basis) appears to be nearly the same as that of Minto.

4.1.9 Trace elements

Table 4.6 gives data on a few of the trace elements: Cl, Hg and V. Our commercial analyst A seems to have gotten suspiciously low values for vanadium. We have nevertheless, for the present, attempted to make vanadium balances (Sect. 12) based on his results.

4.1.10 Particle size analyses

Three lots of Minto coal of different particle size, and one lot of Devco, were used in the combustion trials. Particle size analyses were carried out by sieve analysis. The results are shown in Fig. 4.1 and Table 4.7. As an easily obtainable and useful

Table 4.5. Calorific value of coal (higher heating value), ash-free and moisture-free basis (joules per kilogram of combustible matter).

Coal	Minto ^a	Minto ^b	Devco ^c	Devco ^d
Reporter	Queen's	TM	Queen's	CBDC
Cal. value, MJ/kg	34.6 ± 0.4	34.9 ± 0.3	31.5 ± 0.7	33.54

^a Based on 25 determinations

^b Tibbetts and Montgomery (1976), based on the ten samples

^c Based on 8 determinations

^d Based on data for Devco coal (Prince Mine) furnished to us by Cape Breton Development Corporation

Table 4.6. Trace elements in Minto and Devco coals, dry basis

Coal	Minto ^a	Minto ^b	Devco ^a	Devco ^c
Reporter	Analyst A	TM	Analyst A	CBDC
Cl, ppm	--	--	--	46
Hg, ppm	--	0.37	--	0.05
V, ppm	11	--	7	50

^a Based on three samples

^b Tibbetts and Montgomery (1976), typical run-of-mine sample (ERL no. 2073-76)

^c Data on Devco coal (Prince Mine) furnished to us by Cape Breton Development Corporation, based on product over one week in 1981

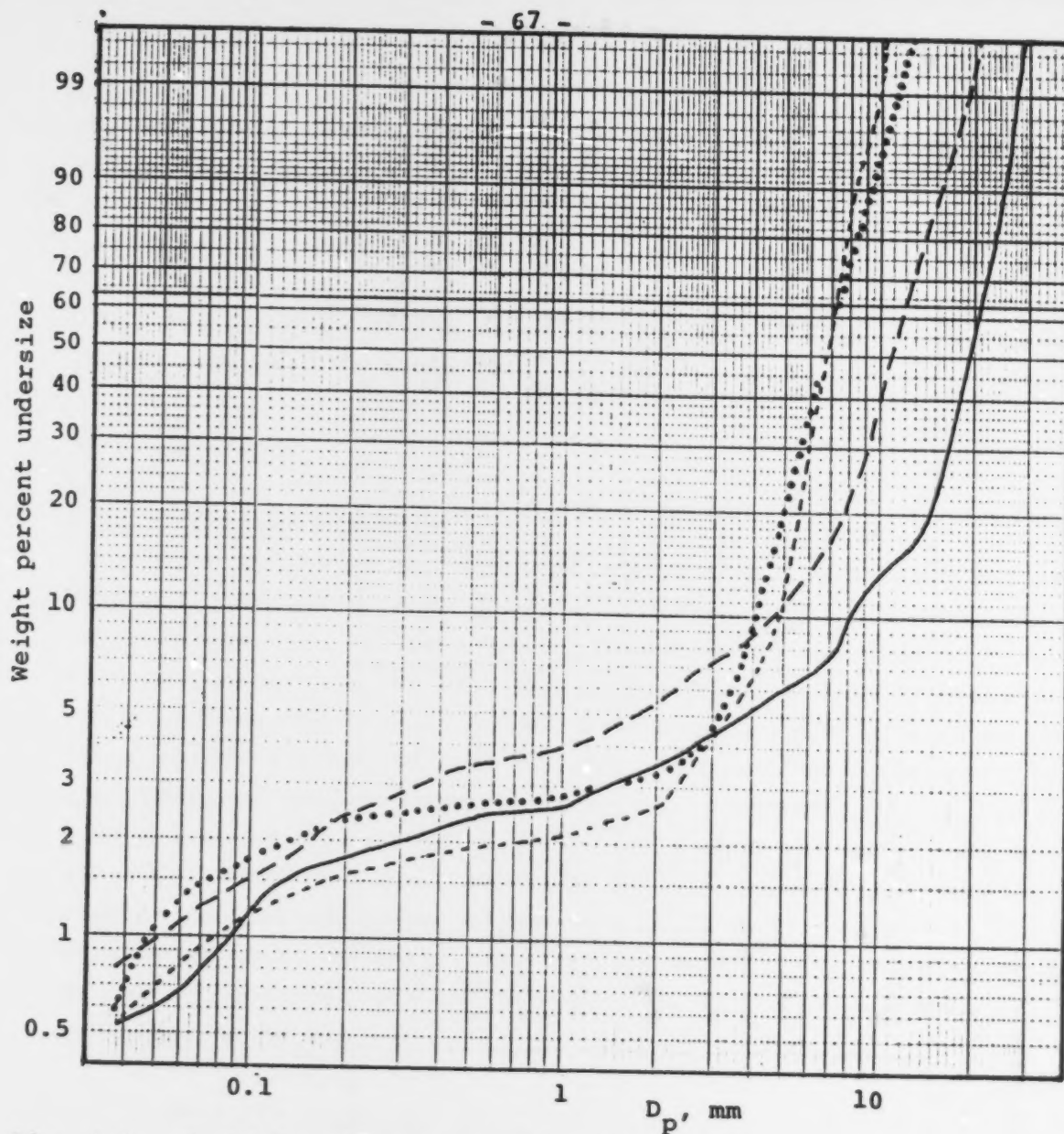


Fig. 4.1. Particle size distribution in the crushed and screened lots of Minto and Devco coal used in the present work.

Key:

Curve	Coal	Mesh (old std.)	$\hat{D}_p$, mm
—————	Minto	- 1 1/2" + 7/8"	19.3
- - - - -	Minto	- 7/8" + 3/8"	10.7
- - - - -	Minto	- 3/8" + 3/16"	6.6
.....	Minto	- 1/4"	6.6

Table 4.7. Sieve analysis of the coals

	Nominal size, as double screened		Fraction undersize, F_D							
			0.01	0.02	0.05	0.1	0.2	0.5	0.8	0.95
Coal	Old, inches	Metric, mm	Particle diameter D_p as determined by sieve, mm							
Minto	-1 1/2+3/4	-38+19	0.085	0.28	3.6	8.0	14.2	19.3	23.1	26.0
	-3/4+3/8	-19+9.5	0.053	0.15	1.6	4.8	7.7	10.7	13.8	17.7
	-3/8+3/16	-9.5+4.7	0.08	0.7	3.3	4.6	5.7	6.6	7.6	8.7
Devco	-1/4	-6.3	0.05	0.13	2.9	4.1	4.9	6.6	9.0	12.0

characteristic particle diameter we have chosen the diameter $\hat{D}_p$ defined by

$$D_p = \hat{D}_p \quad \text{at} \quad F_D = 1/2 ,$$

i.e., the diameter of the sieve opening that passes 50 % by mass of the particles in a sample.

4.2 Limestone

4.2.1 Chemical analysis

Table 4.8 shows the chemical analysis of the present lot of Havelock limestone. Also shown, for comparison, is an analysis of limestone from the same quarry performed at the Mineral Sciences Laboratory, CANMET, Energy Mines and Resources Canada. Assuming that the loss on ignition is all CO_2 , the results on our stone indicate a CO_2 content of 9.67 mol/kg. The element analysis indicates 9.68 mol/kg of Ca and 0.09 mol/kg of Mg, or 9.77 mol/kg of these together, in fairly good agreement with the molar CO_2 content. Since the data on ignition loss have great statistical weight (being based on a large number of determinations), we shall take them as definitive. Lumping Ca and Mg, we accordingly take it that the effective CaCO_3 content of our stone is 96.8 %.

4.2.2 Particle size analysis

Three lots of limestone of different particle size were produced by crushing and screening. A fourth lot consisted of a sized commercial limestone product and was used as purchased. All lots were subjected to sieve analysis to get the particle size distributions. The results are shown in Fig. 4.2 and Table 4.9. We hereinafter use the diameter $\hat{D}_p$, defined as for the coal, as a characteristic particle size.

Table 4.8. Analysis of dry Havelock limestone

Composition by elements			Composition by oxides		
Element	% (QFBC) ^a	% (MSL) ^b	Oxide	% (QFBC) ^a	% (MSL) ^b
Al	0.50	0.08	Al ₂ O ₃	0.94	0.15
Ca	38.8	39.52	CaO	54.3	55.33
Fe	0.37	0.12	Fe ₂ O ₃	0.53	0.17
Mg	0.22	0.41	MgO	0.36	0.68
Si	0.44	0.63	SiO ₂	0.94	1.35
S	--	< 0.01	SO ₃	--	< 0.03
Total oxides				57.07	57.68
Loss on ignition at 980 °C				42.57 ± 0.65	42.60
Oxides plus ignition loss				99.64	100.28
Other MSL results:					
Est. Ca present as CaO + Ca(OH) ₂ :			0.05 %		
Est. Mg present as MgO + Mg(OH) ₂ :			< 0.01 %		
Est. Ca present as CaSO ₄ :			< 0.01 %		

^a Present lot of stone. Element analyses by analyst A. Loss on ignition obtained in our laboratory (mean in scores of runs in our bench-scale studies of calcination kinetics)

^b Analysis by Mineral Sciences Laboratory, CANMET, Energy Mines and Resources Canada (their sample no. MS-CL-78-327)

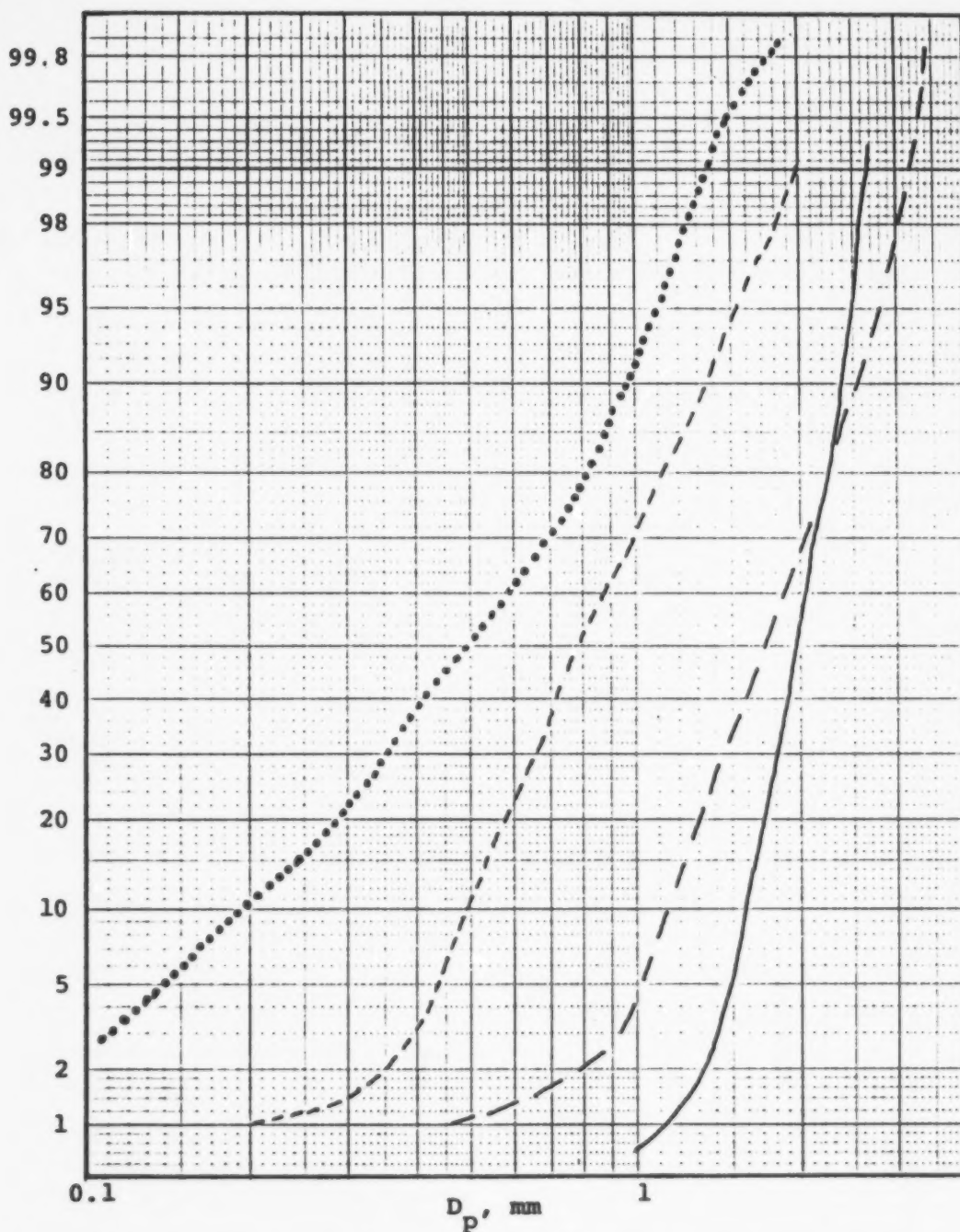


Fig. 4.2. Particle size distribution in the crushed and screened lots of Havelock limestone used in the present work.

Key:

Curve	Mesh (old std.)	$\hat{D}_p$, mm
—————	- 1/8" + 11	1.98
- - - - -	- 6 + 20	1.72
- . - . - .	- 11 + 34	0.78
.....	- 16 + 100	0.49

Table 4.9. Limestone particle size distributions

Nominal size as		Mass fraction undersize, F_D								
double screened		0.01	0.02	0.05	0.1	0.2	0.5	0.8	0.95	
Old, mesh	Metric, mm	Particle diameter D_p as defined by sieve analysis, mm								
-1/8+11	-3.1+1.8	1.13	1.33	1.48	1.59	1.72	1.98	2.23	2.47	
-6+20	-3.3+0.9	0.45	0.79	1.01	1.14	1.31	1.72	2.27	2.80	
-11+30	-1.8+0.6	0.20	0.35	0.44	0.49	0.58	0.78	1.12	1.34	
-6+100	-3.3+0.15	0.035	0.083	0.141	0.197	0.29	0.49	0.96	1.08	

5. APPROACH TO STEADY STATE, AND OTHER TIME-DEPENDENT ASPECTS OF BEHAVIOUR

5.1 Introduction

Various types of time-dependent behaviour were observed, with widely differing time constants. Thus, before turning to the results for "steady-state performance", which is in fact only statistically stationary, it is necessary to examine the forms and implications of these time-dependent components of system response.

5.2 Phenomena with the smallest time constants

The fastest variations of variables occur in the fluctuations of local temperature, composition and velocity that are due to the strong agitation prevailing in a well-fluidized bed (as with the passage of bubbles and jets, surging of the emulsion, etc.). The time scale for these fluctuations is typically on the order of seconds or fractions of seconds. The time constants of our measurement systems were all too great to let them follow such fluctuations. For some variables, there are available other systems that do have fast enough response, e.g., fine--and unfortunately fragile--thermocouple probes for temperature and zirconia probes for oxygen potential. Such measurements, though, were of low priority for the present objectives and were not undertaken.

5.3 Phenomena with time constants typically 10--100 s

Response to changes in many of the principal operating parameters fell within this range. For example, a step change in coal or limestone feedrate, perhaps due to jamming of a feeder or the airlock valve, generally produced visible response within 10 s. Changes occurring on these time scales were within the capabilities of our measurement systems and were closely followed.

The one outstanding, and readily explainable, time-dependent behaviour of this class occurring at fixed settings of the system controls was a fluctuation in freeboard gas temperatures and concentrations ascribable to two basic types of variation in coal feedrate. Both types were accentuated by low feedrate. The first, illustrated by Fig. 5.1, was wholly correlated with the turning of the feeder helix when feeding fine (small-particle) coal, producing a regularly fluctuating delivery. The combustor itself was clearly able to respond, no doubt mainly through the coal devolatilization and subsequent volatiles burning. The second, illustrated by Fig. 5.2, supercedes the first when very large coal particles are slowly fed; then lumps leave the feeder virtually one at a time, in a train that is irregular in both lump size and spacing, producing highly random fluctuations in freeboard temperature and concentration, again because of fluctuating volatiles production. In a system large relative to our pilot plant, neither of these effects might be perceptible.

Although interesting by themselves for what they tell about certain aspects of combustor response, these fluctuations are of minor importance to the overall analysis of our results (e.g., in the determination of the steady state performance), in which they figure largely as a form of noise. We have therefore, for most further data analysis, filtered them out by time-averaging:

$$\langle f(t) \rangle_{\Delta t} = \frac{1}{\Delta t} \int_{t-\Delta t}^t f(t) dt, \quad (5.4.1)$$

where Δt is a suitable averaging period. This averaging was done by the on-line computer, normally with $\Delta t = 60$ s. We will not introduce any special notation for the so-smoothed data, but shall take it as understood hereinafter that all values of temperature, pressure and gas concentration are of this kind unless otherwise noted.

RUN 830316 GAS PROBE @PORT 91 TIME=11.7 Hrs.

Full Scale Values:
 H₂CO 1 % B:SO₂ .2 % C:CO₂ 20 % D:O₂ 25 % E:NO 1000 PPM F:HC 262 PPM

COAL: -1/4". LIMESTONE: -16 + 170 MESH

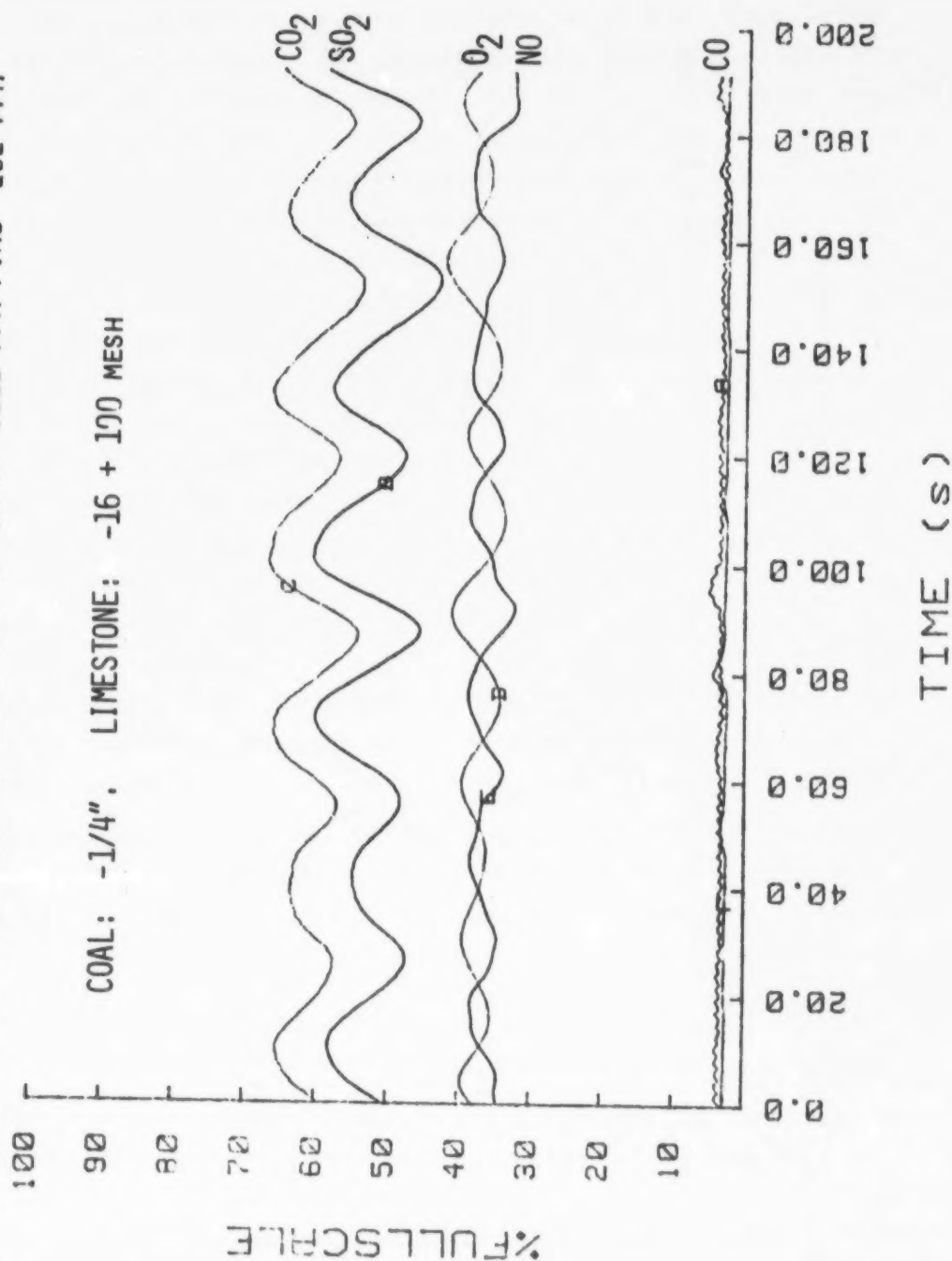


Fig. 5.1. Fast fluctuations in combustor exit gas composition due to fluctuating coal feed rate when feeding fine coal particles.

RUN 300330 GAS PROBE @PORT 69 TIME=9.65 Hrs.

Full Scale Values: --- HC measured separately
 R:CO 1% F:CO2 12% C:CO2 22% F:O2 5% X:NO 1200PPM H:HC 10482PPM

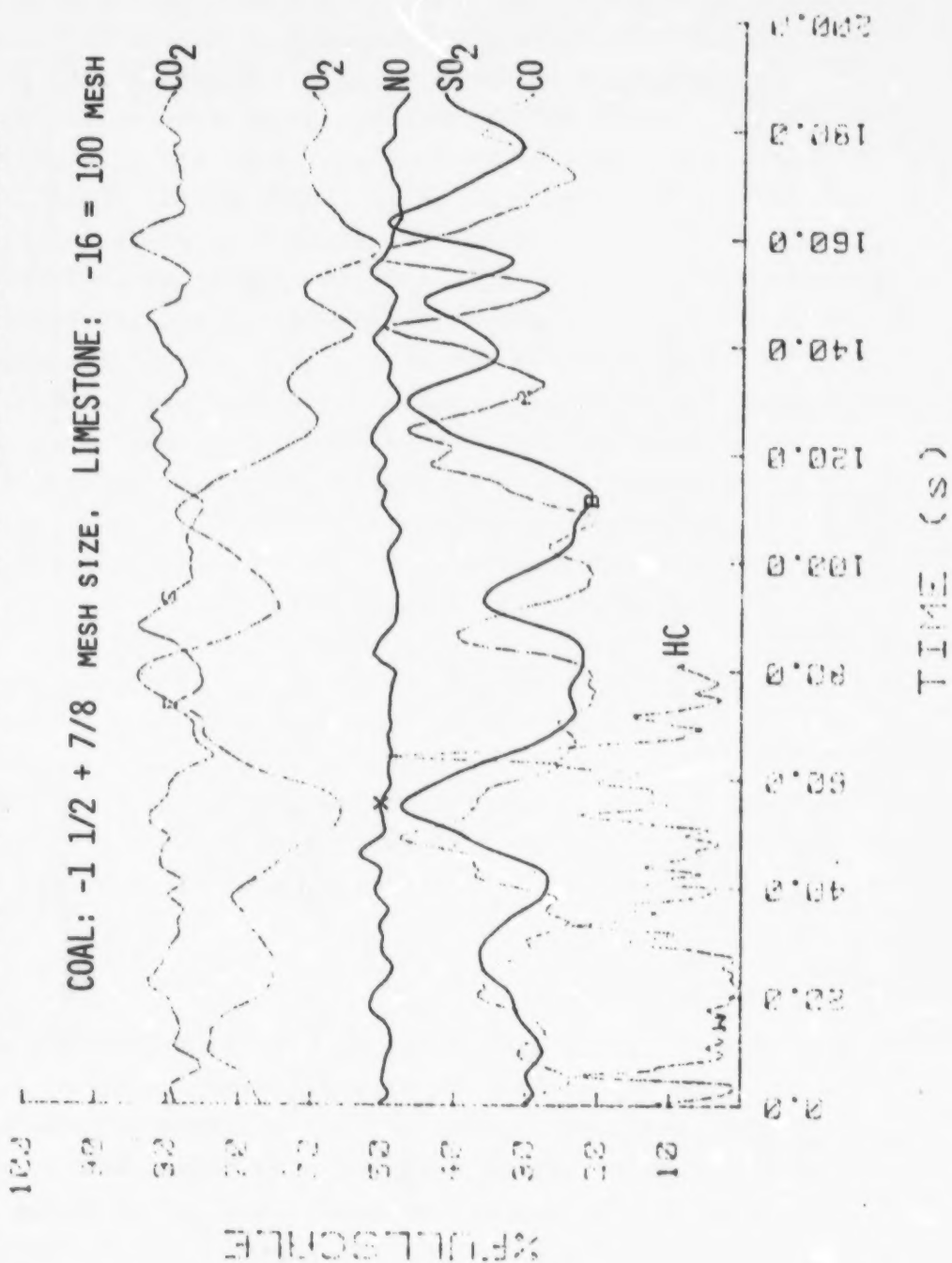


Fig. 5.2. Fast fluctuations in combustor exit gas composition due to fluctuating coal feed rate when feeding coarse coal particles.

5.4 The approach to steady state

Figure 5.3 shows an example of records of bed temperature and of gas composition at combustor exit as provided by the on-line computer. There is evidence of approach to steady state in most of these records, excepting those of NO and O₂. Nitrogen oxides emission almost always tended to be virtually constant throughout a run, as soon as the bed temperature rose above 800 °C. Of course, when in-run adjustments in coal feedrate were made--normally only in the very early part of a run--the emission of nitrogen oxides changed too. The coal feedrate over the records shown in Fig. 5.3, and in other illustrations in this Section, was fixed. The oxygen record in Fig. 5.3 is fairly typical of runs in which the excess air level was low; the O₂ concentration was then sensitive to small changes in coal feedrate (due to inconsistencies in the coal, changing level in the coal feed hopper, etc.) and tended to drift about enough so any systematic changes with time were below this "noise" level.

The CO₂ record is not shown in Fig. 5.3, nor in any other figures hereinafter, because at given air, limestone and coal feedrates, the sum $X_{O_2} + X_{CO_2} + 0.5 X_{CO}$ was virtually a constant for all positions and times in the freeboard, where X_i is the mole ratio of gas component i, defined as moles of i present per mole of water-free gas. Further, because X_{CO} was generally small, $X_{O_2} + X_{CO_2}$ was virtually a constant. In the run to which Fig. 5.3 pertains, $100(X_{O_2} + X_{CO_2} + 0.5 X_{CO}) = 19.41 \pm 0.08$, and $100(X_{O_2} + X_{CO_2}) = 19.26 \pm 0.09$. These results, of course, (a) simply reflect element conservation for oxygen, and (b) indicate high accuracy of the gas analysis for O₂ and CO₂.

The overall apparent residence time of bed solids, $\tau_p = \dot{m}_b / \dot{m}_p$, was about 2.3 hr for the run in Fig. 5.3, and the (likely more significant) residence time of bed product, $\tau_b = \dot{m}_b / \dot{m}_b$, was 5.8 hr. In Fig. 5.3 the temperature appears to level off at about 5 hr COT, the SO₂ at around 3-1/2 hr, unburnt hydrocarbons

COAL: -1 1/2 + 7/8 MESH

LIMESTONE: -16 + 100 MESH

FBC2 BED MONITORING

830330

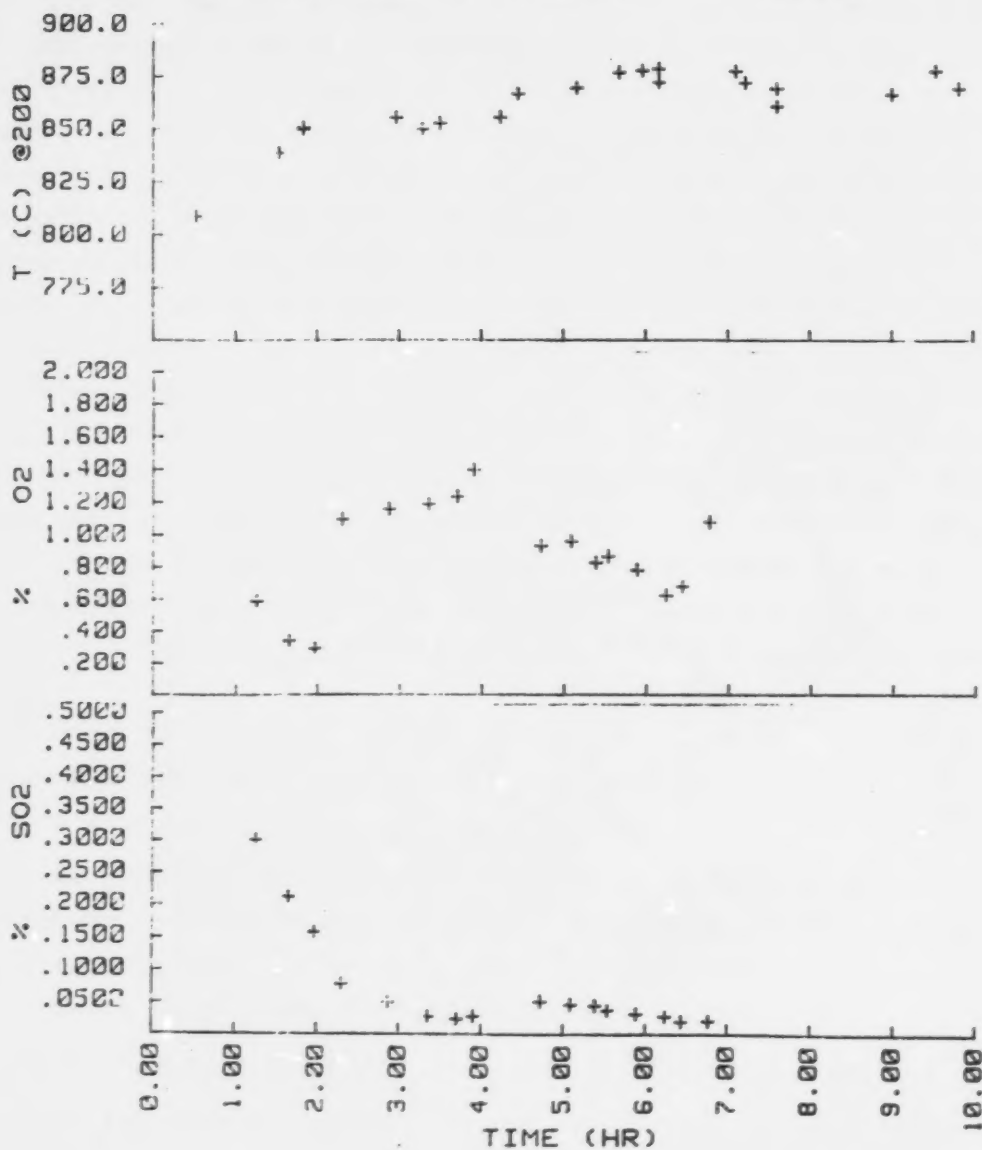
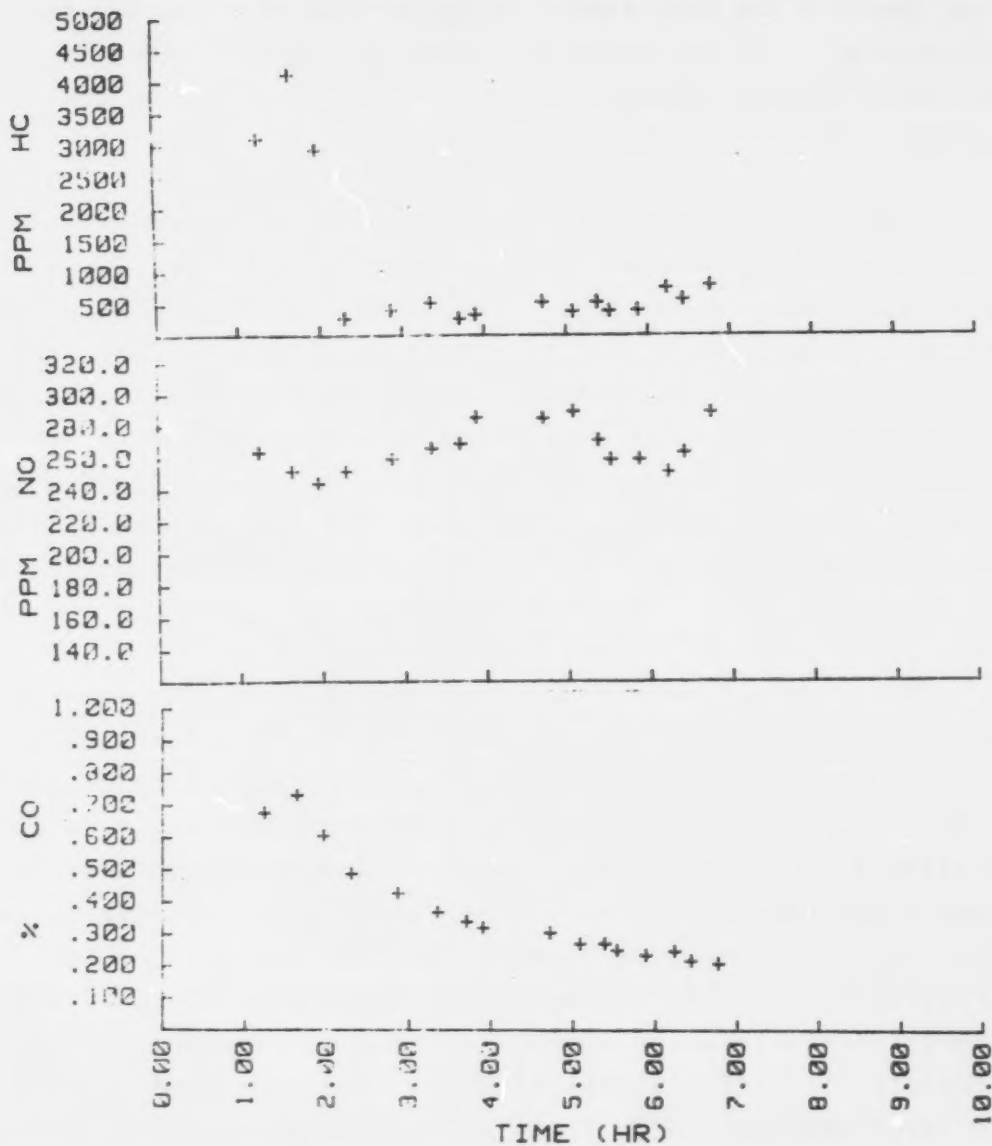


Fig. 5.3. Examples of records of bed temperature and of gas composition at combustor outlet over the duration of a run (run 830330).

P91

GAS CONCENTRATION 830330



at around 2-1/2 hr, and CO at around 6 hr. Gas compositions and bed temperature thus appear to be virtually at steady state after passage of just one residence time τ_b . This finding is indeed a fairly general observation, valid for most of the trials in the present work.

Most of the records in Fig. 5.3 show a component of slow, seemingly random variation. Bed temperature, for example, in the virtual "steady state" after 5 hr COT fluctuated over a 20 °C range. These slow fluctuations are discussed in Sect. 5.6, but for the moment we shall simply treat them as a form of noise which makes the steady state, and thus the approach to that state, more difficult to recognize and quantify. Basically, a statistical approach should be used, but limitations of time and funding did not permit repetition of runs or great prolongation of runs just to get better statistics.

Gas composition and bed temperature give some indication of the approach of steady state, but they are not decisive alone. Solids compositions may be expected to be much more critical. If one starts a run with a bed consisting of material significantly different from the eventual steady-state material, then it is surely necessary to flush out most of the initial charge before the bed composition can closely approach its final form. The experimental results indeed support this expectation. Figure 5.4 shows records on a run which started with a relatively mild disequilibrium (hot start, with the limestone feedrate decreased to lower the Ca/S ratio from 3.52 in the preceding run to 2.27, but with some disturbance in the first hour because of a passing problem with the recycle system). Carbon analyses were obtained on all solids samples taken after the fifth hour. The figure shows, as a fraction of the carbon fed in the coal, the carbon fluxes carried by the cyclone product and the bed product: $\dot{m}_{C,c}/\dot{m}_{C,f}$ and $\dot{m}_{C,b}/\dot{m}_{C,f}$. It appears from these results that the steady state is approached in the expected asymptotic manner, and is virtually attained at 13 hr run time. Here the

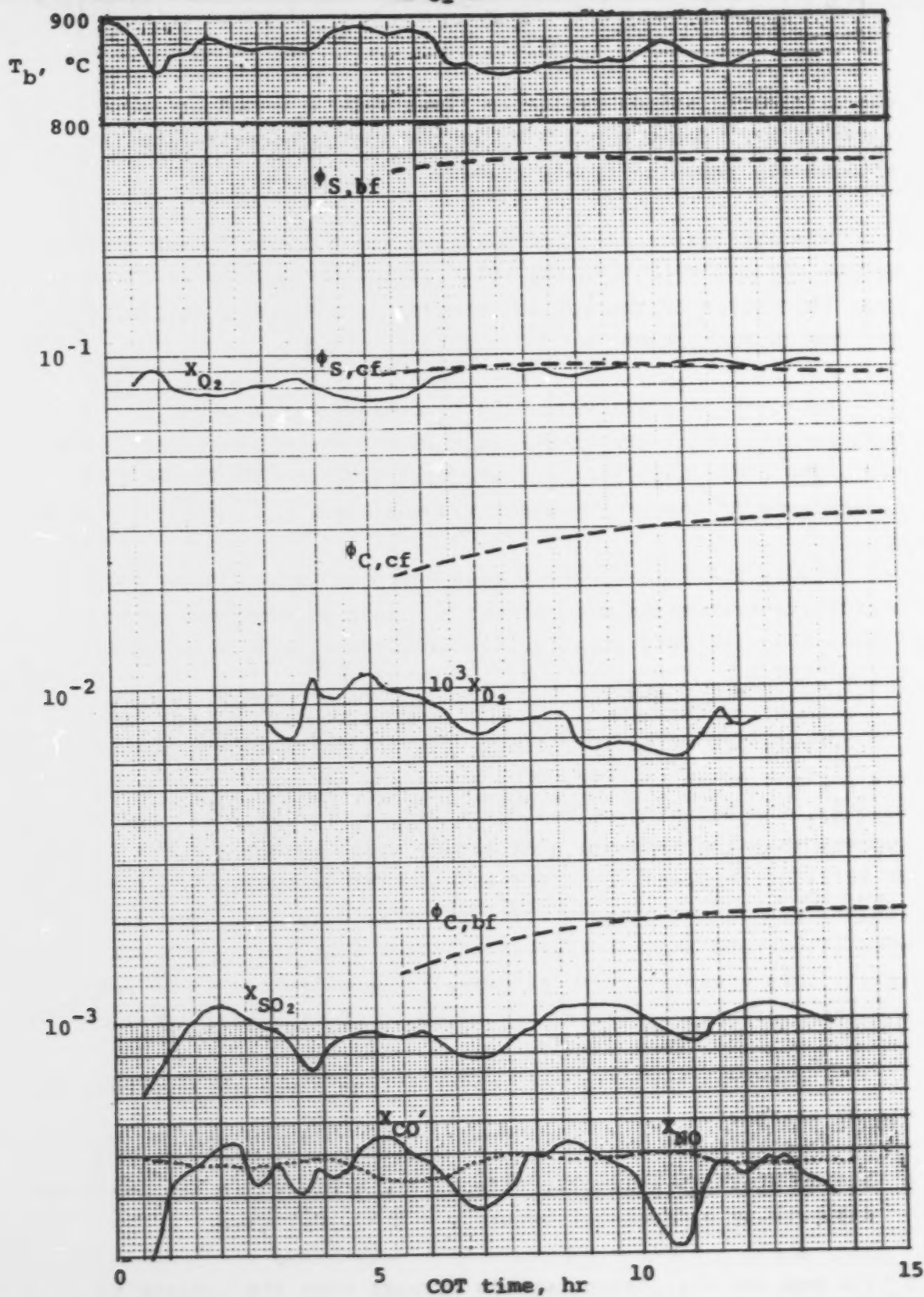


Fig. 5.4. Characteristics of gas and solids streams leaving the combustor during run 830316.

Note: $X_{CO_2} + X_{O_2} = 21.55$

characteristic solids residence times were $\tau_p = 4.8$ hr and $\tau_b = 7.6$ hr. Hence $t \geq 1.7 \tau_b$ is the indicated criterion for effectively reaching steady state.

Similar indications are given by the sulphur fluxes carried by the solids, Fig. 5.4; the total variation in sulphur fluxes over the course of the run was however considerably smaller than for the carbon fluxes.

Among the gas analyses, the O_2 concentration gives a noticeable hint of the change in carbon content occurring in the bed. The other analyses indicate virtually nothing above the noise level of the usual random fluctuations.

Figure 5.5 illustrates a case where sulphur flux of the solids streams varied more strongly. Both sulphur and carbon fluxes here indicate an effective equilibration time around 12 hr, or $1.9 \tau_b$.

Figure 5.6, finally, illustrates a case where the results are unclear about the effective beginning of steady state. However, the bed-product residence time τ_b was 10.8 hr, and it appears that the assumption of steady state beyond $t = 2 \tau_b = 22$ hr is quite plausible. These data, incidentally, are for the longest run in the present work, of 30 hr duration. The data also illustrate another feature of cases with low excess air level; this usually leads, as here, to large swings in CO and unburnt hydrocarbon levels as well as in the O_2 level itself.

In summary, our principal observations about the approach to steady state are:

- o steady state is evidently effectively achieved in all important respects by or somewhat before $t = 2 \tau_b$, COT time
- o the bed and the leaving solids streams take the longest to

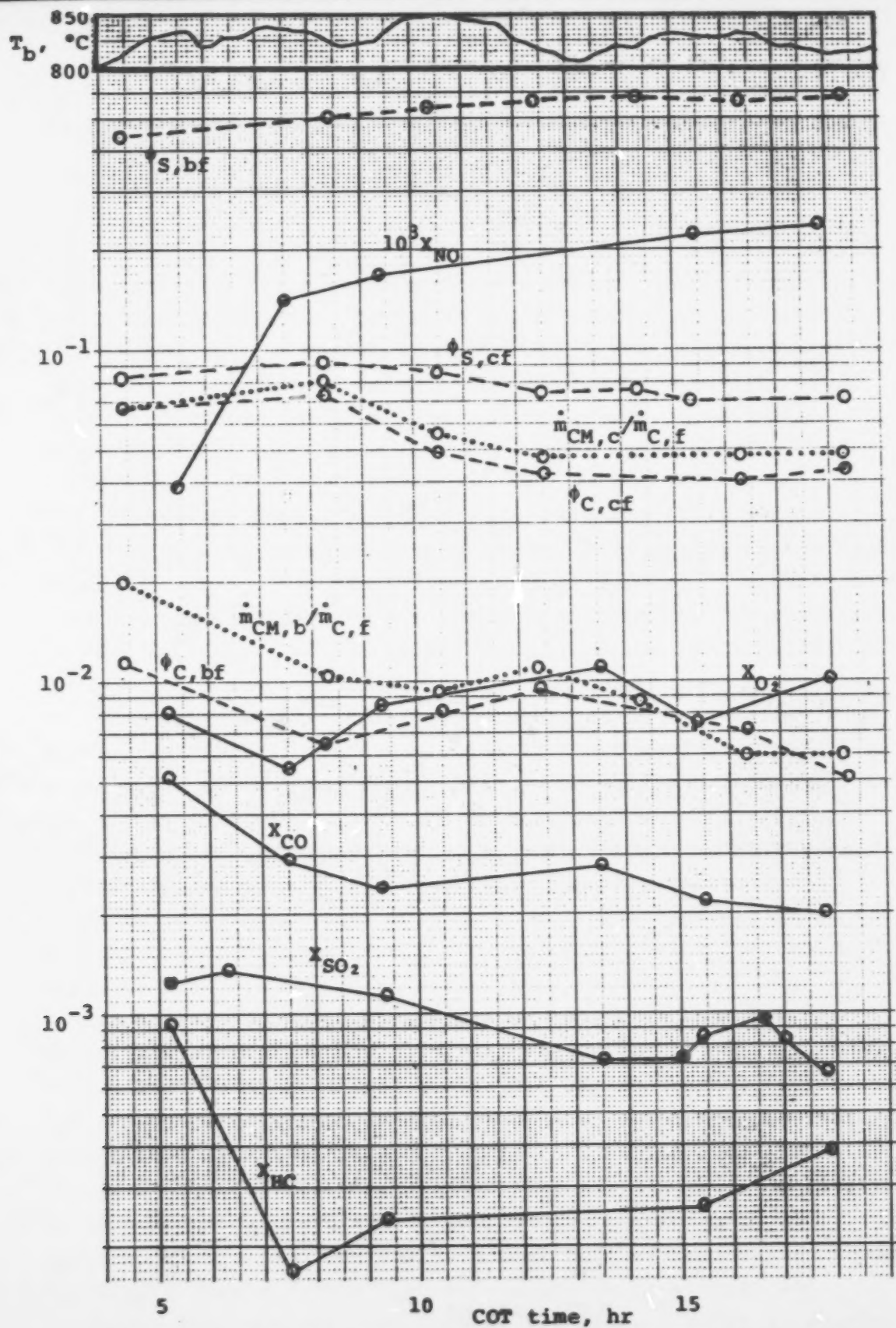


Fig. 5.5. Characteristics of gas and solids streams leaving the combustor during run 838281.

Note: $X_{\text{CO}_2} + X_{\text{O}_2} = 18.65$

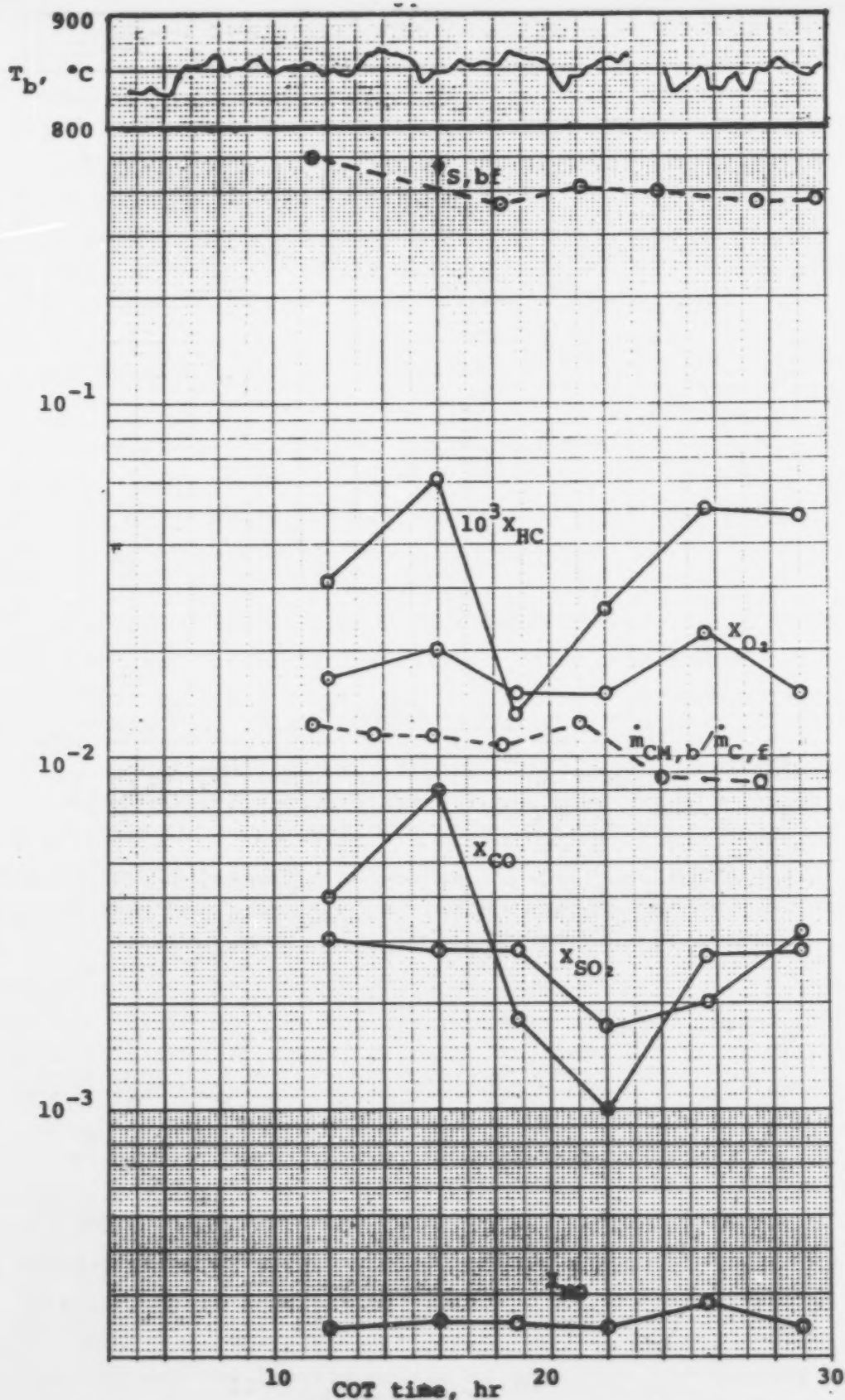


Fig. 5.6. Characteristics of gas and solids streams leaving the combustor during run 821116.

Note: $X_{\text{CO}} + X_{\text{O}_2} = 18.48$

reach steady state. Gas compositions and bed temperature, however, usually settle by $t = \tau_b$ or even sooner.

- o in general, it is recommended that $t \geq 2 \tau_b$ be adhered to as the criterion of steady state. Indications from in-run monitoring, such as the levelling off of temperature and SO_2 emission, can be seriously misleading. Very mistaken results may be obtained, for example, for the combustion efficiency which depends heavily on the carbon content of the solid products and takes significantly longer to settle.

5.5 Slow fluctuations or drift

The slow fluctuations noted in Sec. 5.4, occurring throughout a run and perhaps best described as erratic drift, typically have time constants ranging from 10 minutes for the fastest to 3 hr for the slowest. The possible explanations are of two kinds:

1. The fluctuations reflect a form of system instability. Reactor instabilities are possible (Aris and Amundsen 1958), for example, in which a reaction system (with an exothermic reaction and a cooling load) has multiple steady states for fixed input conditions and erratically jumps between these. In this vein, the records of some of the present runs seem to show, for instance, bed temperature staying at a nearly constant level for several hours, with only small fluctuation, then quite rapidly going to a new level and hovering there for a while, only to change again. Multiple steady states can exist even for relatively simple systems. The present system is complex enough, with a mixture of important exothermic and endothermic reactions (coal pyrolysis, pyrolysis gas combustion, heterogeneous char combustion, limestone calcination and sulphation), so multiple steady states and attendant instability are not at all unlikely. It is noteworthy that in some very early runs in which we operated with a sand bed and no limestone, bed temperature tended to be very stable. This stability question is a large and difficult one, however, and outside the present scope to pursue.

2. The fluctuations simply reflect the effects of variations in the input parameters, primarily the feedrates of coal, limestone and air. Here fluctuation of airflow can be dismissed immediately, since this flow was extremely steady and varied negligibly relative to the coal and limestone. The coal and stone flows, however, may indeed have varied enough to explain a substantial part, or even all, of the observed effects. The variations in bed temperature, for example, spanned a range as small as 30 °C and as large as 60 °C. A 30 °C temperature variation superimposed on a rise from 20 °C to 850 °C only amounts to a 4 % effect. This could easily be explained by a fluctuation in coal feedrate with standard deviation under 2 %, a feeder performance that is quite acceptable by any reasonable standard. Under the most favourable conditions--when fed with small dry particles of coal or limestone--our feeders seemed to be capable of constant delivery within ± 1 %. There was an apparent tendency in some runs for the feedrate of coal to change slightly with level in the feeder hopper. It also varied with coal properties such as moisture content and fines content which were not quite uniform from one barrel lot to the next (both coal and limestone were stored and handled in steel barrels).

An indication of typical coal feeder performance as gauged by the weight record of the coal feeder hopper is given by Fig. 5.7. The coal feedrate control here was fixed after 4.5 hr COT. Analysis of the data after 4.5 hr gives a coal feedrate of 29.5 ± 0.9 kg/hr, or a constancy of feedrate within ± 3.1 % standard deviation. There is some indication that the feedrate tended to be lowest when the feeder hopper was full and increased slightly as the bin emptied. There is no apparent correlation, however, between this tendency and the fluctuations in bed temperature. The coal in this example, it should be noted, was of the largest particle size used and the most difficult to feed at uniform rate.

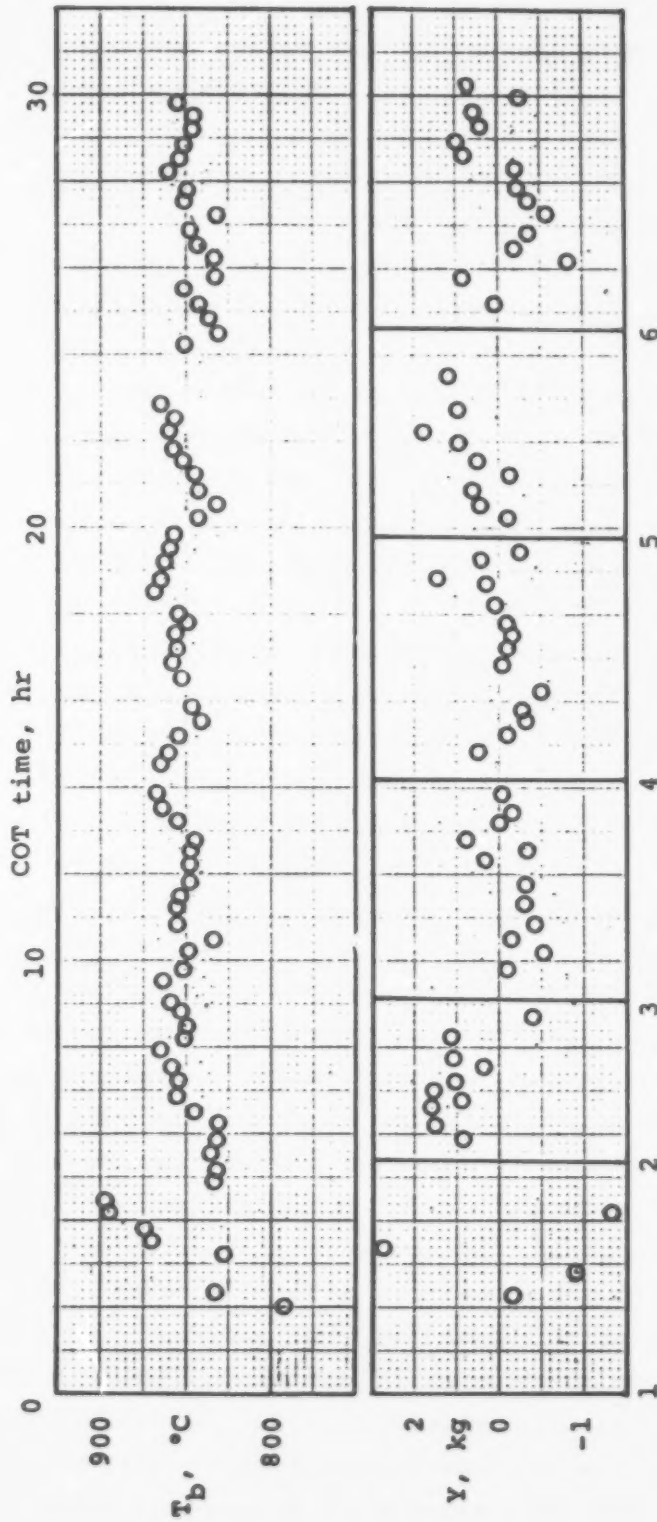


Fig. 5.7. Illustration of the degree of constancy of coal feedrate (run 821116). The ordinate on the lower graph is $Y = (m_n - m) - \langle \dot{m} \rangle (t - t_n)$, kg. The time-average coal feedrate $\langle \dot{m} \rangle$ is 29.5 kg/hr. The calculation had to be restarted after each recharging of the coal feed hopper; m_n is the mass of coal in the hopper at time t_n , the start of n 'th calculation period.

A further critical examination of the implications of slow fluctuations is given in Sect. 13.

6. FREEBOARD VARIATIONS IN TEMPERATURE AND CONCENTRATION

6.1 Introduction

In this Section attention is turned from temporal variation of temperature and concentration at given points to the question of spatial variation at a given time. The time of interest, however, is simply the steady state. It was seen in Sect. 5 that in the steady state, as before it, there exists a certain complication from random fluctuations of the temperature and concentration at any point in the freeboard (and presumably in the bed too). This problem could be treated statistically by time or ensemble averaging if runs were sufficiently protracted or many times repeated. Since we could not afford great protraction, and no repetition, attention is particularly focussed in this section on periods effectively within the "steady state" (or statistically stationary state) and when bed temperature varied negligibly for at least 60 minutes while spatial traverses with gas sampling probes were made. However, because this condition could only be satisfied by chance, it was not always fully met in the examples here presented.

There were problems with the freeboard gas sampling probe which were not satisfactorily solved until over half the runs had been completed. These problems stemmed from the excessively small size of the probe which was required because of a wrong (as it turned out) design decision in sizing the combustor probe ports: measuring 21 mm I.D., they were too small for a comfortably sized cooled probe. As a result, freeboard traverses of gas composition were obtained only for the later runs. As explained in Sect. 6.2, we are planning to remove the freeboard heat exchanger. This will open space to insert a conveniently larger sampling probe downwards from the top which will be freely positionable vertically to obtain traverses of the freeboard. It will also facilitate isokinetic sampling.

We used a separate large gas sampling probe at the combustor exit gas sampling port (which was much larger than the others), and this never gave any difficulty. The main problem with the small traversing probe was with the dust collection and removal, accomplished with a thimble-shaped ceramic filter.

6.2 Vertical temperature profiles

Figure 6.1 shows a typical vertical temperature profile in the combustor as indicated by the fixed thermocouples. The indicated temperature falls slightly until the freeboard heat exchanger is neared, and then it drops quickly. The figure clearly illustrates a flaw in the combustor design: omission of the freeboard heat exchanger would provide an additional metre of clear freeboard reaction space and avoid the quenching, cooling action of this heat exchanger on the approaching dust-laden gas and the adjacent combustor walls. We are therefore planning to remove this exchanger and replace it by an additional outboard unit in future work. Laboratory ceiling height limits further upward extension of the combustor.

The freeboard temperatures indicated by the fixed thermocouples are useful, but they cannot reliably be interpreted as gas temperatures. The probes were of the sheathed type, of 9.5 mm sheath diameter, and obviously suffered greatly from radiation error. An aspirating thermocouple probe has been built and will be used to obtain freeboard gas temperatures in future work.

6.3 Freeboard gas composition profiles: NO

Figure 6.2 shows four NO concentration profiles, representative of runs in which NO dropped quite markedly with height. This freeboard removal of NO has been shown (Walsh, et al. 1982) to be due to heterogeneous reaction with elutriating char, and is a common occurrence. Our freeboard was unfortunately often not high enough (nor was that of Walsh, et

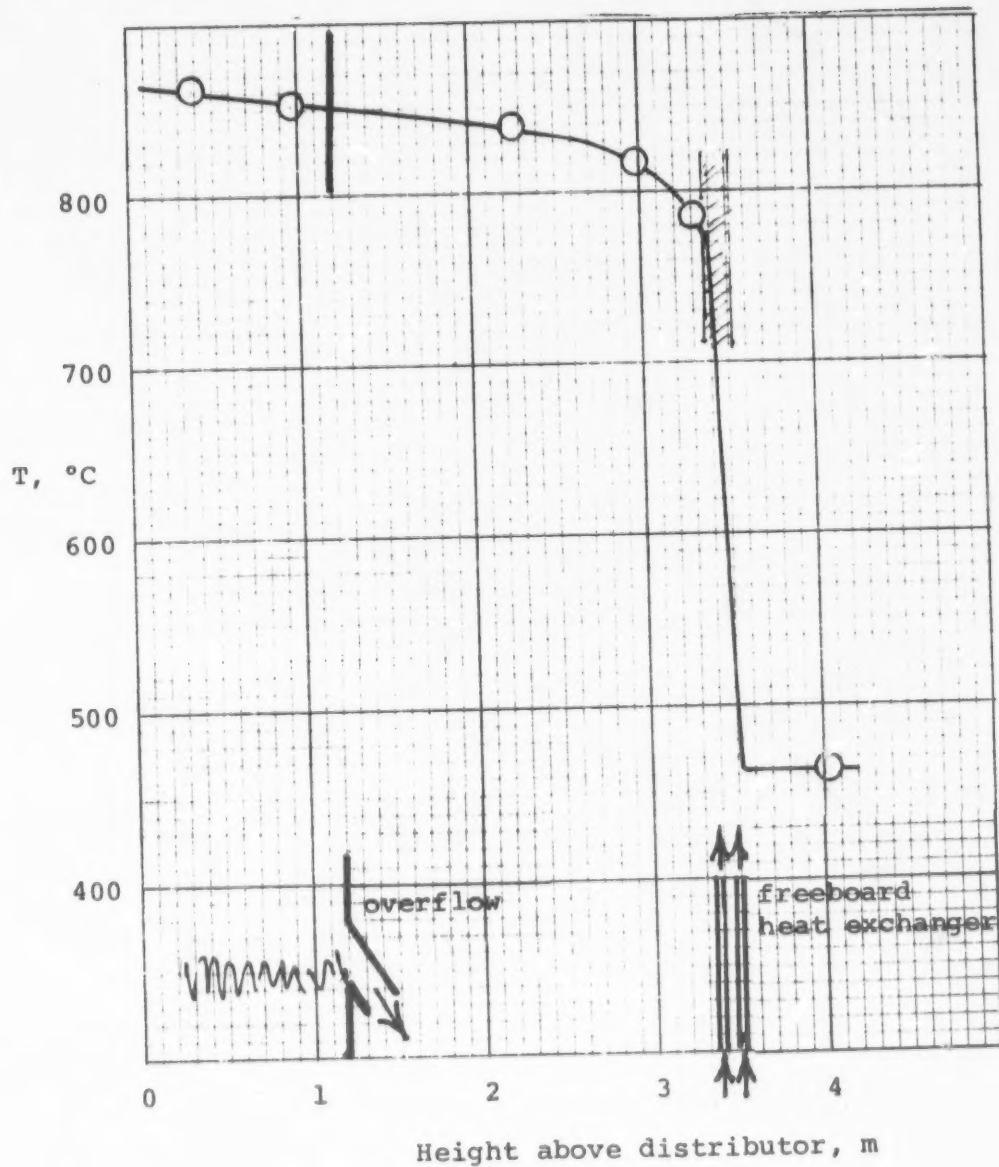


Fig. 6.1. Vertical variation of temperature as indicated by the fixed thermocouples, run 830324.

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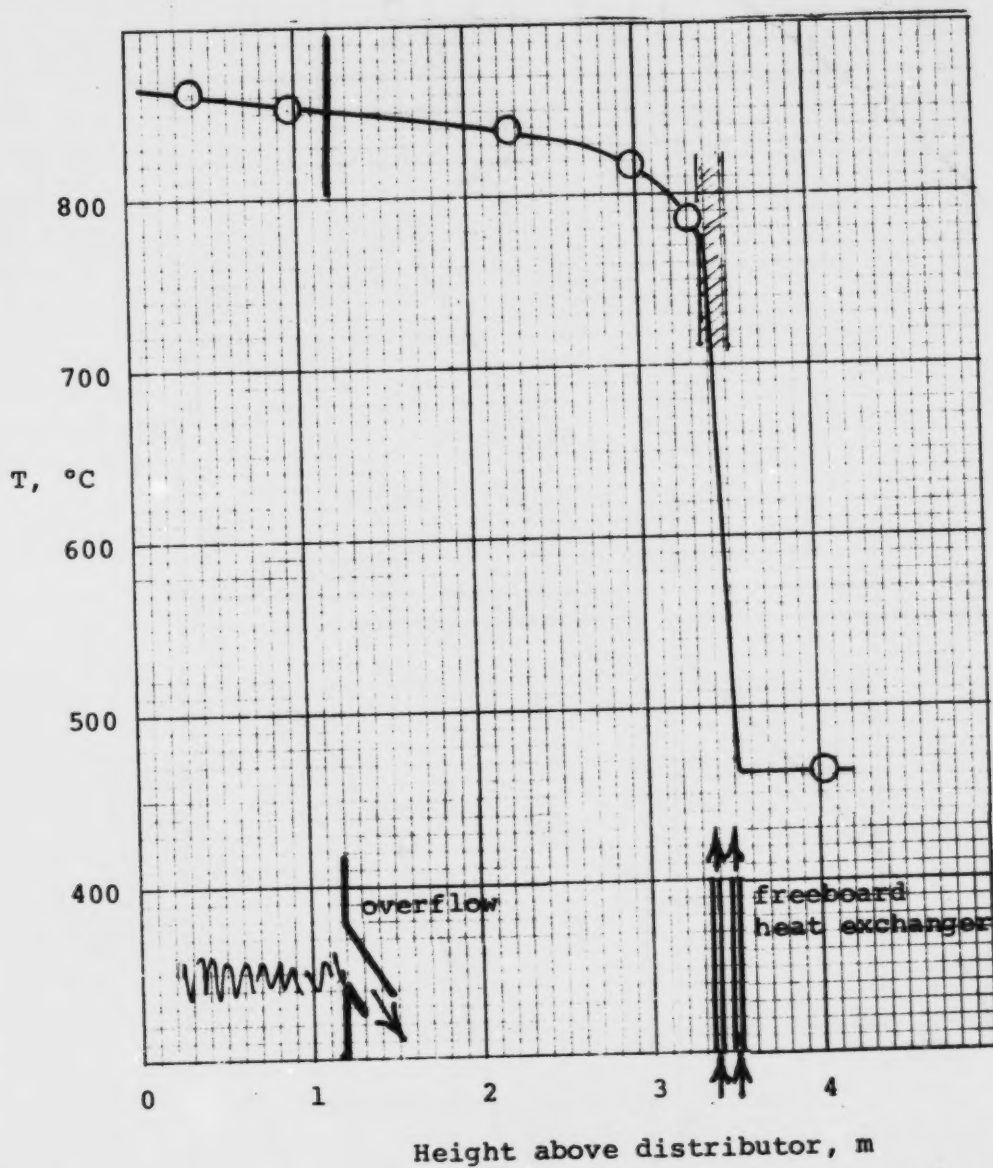


Fig. 6.1. Vertical variation of temperature as indicated by the fixed thermocouples, run 830324.

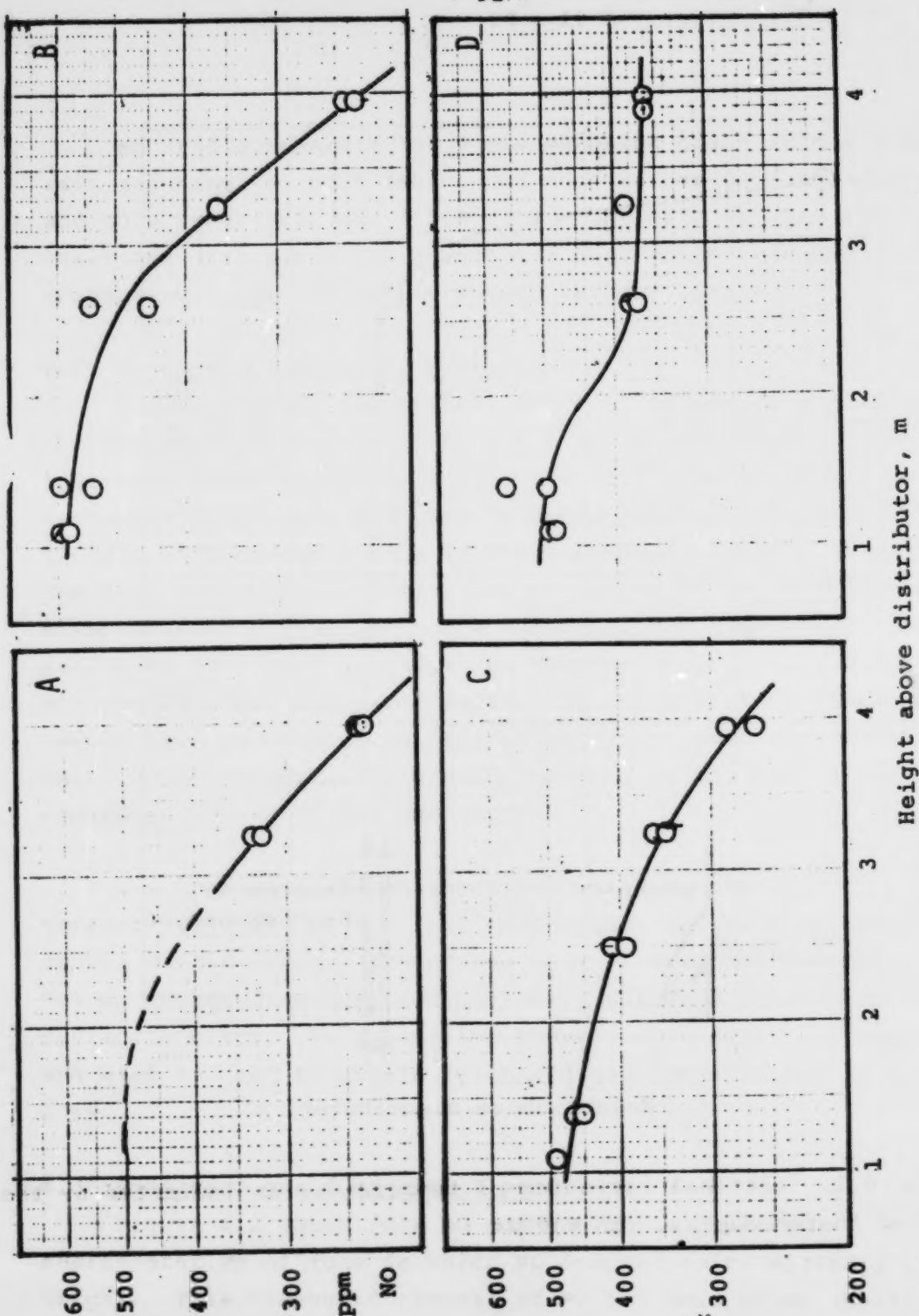


Fig. 6.2. Freeboard profiles of NO content of gases, ppm by volume, dry basis. See Table 1.1 for operating conditions.

Graph

A

B

C

D

Run

830322

830324

830330

830329

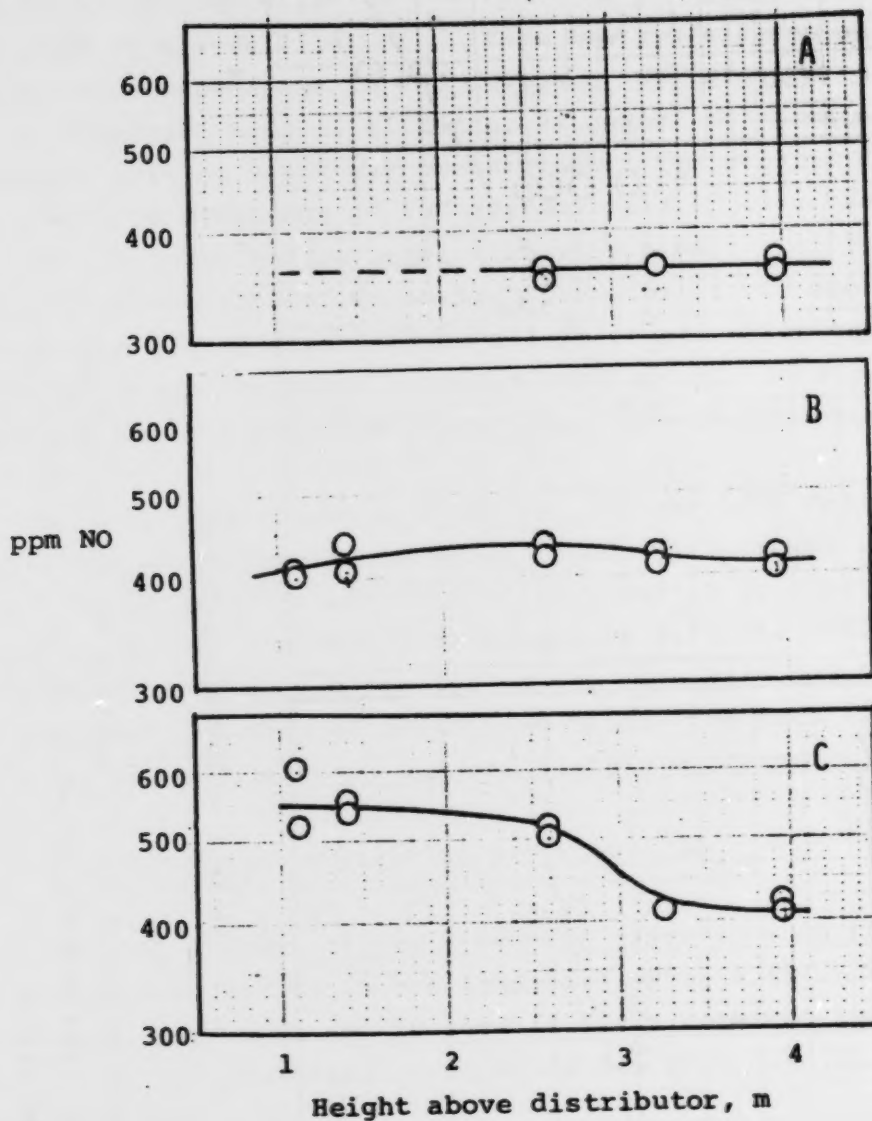


Fig. 6.3. Freeboard profiles of NO content of gases, ppm by volume, dry basis. See Table 1.1 for operating conditions.

Graph	A	B	C
Run	830316	830315	830309

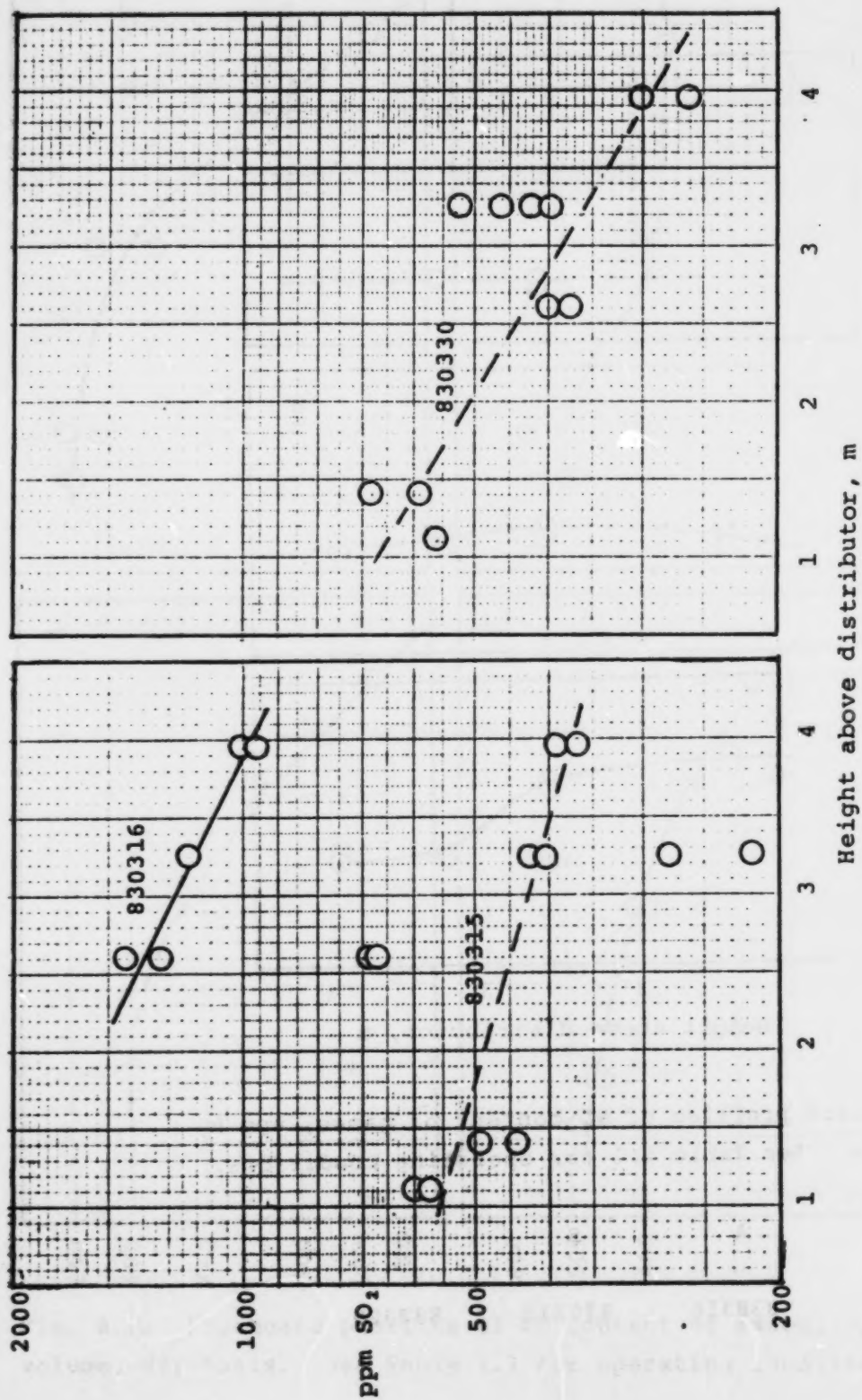


Fig. 6.4. Freeboard profiles of SO₂ content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

al.) to allow completion of the reaction to equilibrium (as indicated by a levelling off). This feature exposes an element of apparatus dependence in FBC combustor performance; depending on the freeboard height, different values of NO emission at combustor exit or stack may be obtained. The problem can be well dealt with by designers of FBC systems only if they have available a quantitative model that will predict the freeboard NO removal. For our part, we would be happier if our combustor had a very much higher freeboard so all freeboard reactions could be followed to any stage and to their conclusion. As already remarked, we lack laboratory space for such an expansion.

Not all runs showed such marked freeboard NO reduction. Figure 6.3 shows results on three in which there was little or none. These were all with Devco coal, and it is likely significant that the mass flux of carbon with the cycloned elutriated material was smaller in the runs with Devco coal than in the great majority of those with Minto (the flux in the baghouse product, though, was not visibly different).

6.4 Freeboard gas composition profiles: SO_2

The SO_2 content of the freeboard gases in some runs decreased with height in the freeboard, Fig. 6.4; in some it increased, Fig. 6.5; and in some it first rose and then fell, Fig. 6.6. The processes responsible for this complex behaviour evidently are:

- o freeboard burning of char, producing SO_2
- o freeboard capture of SO_2 by sulphation of CaO
- o freeboard decomposition of CaSO_4 , producing SO_2
- o freeboard adsorption or desorption of SO_2 by solid particles, taking from or adding to the SO_2 content of the gases.

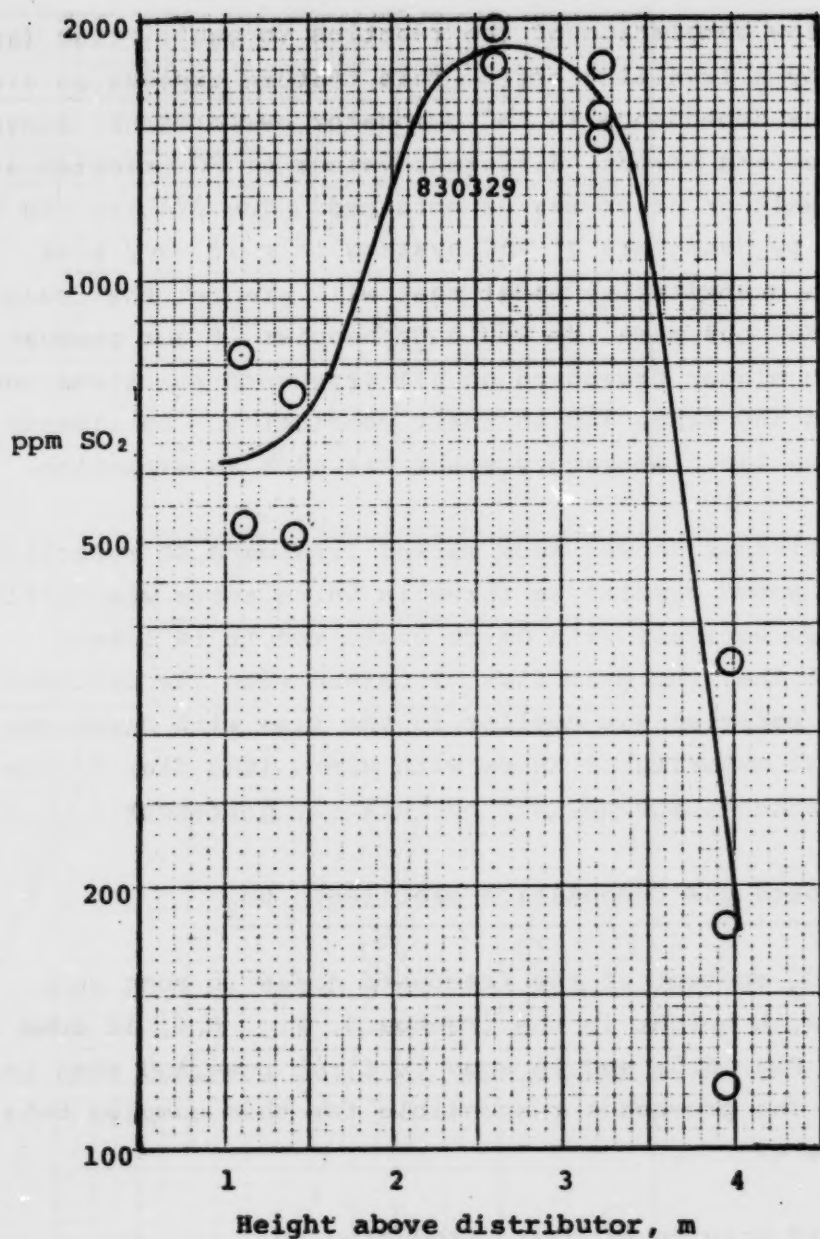


Fig. 6.6. Freeboard profiles of SO_2 content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

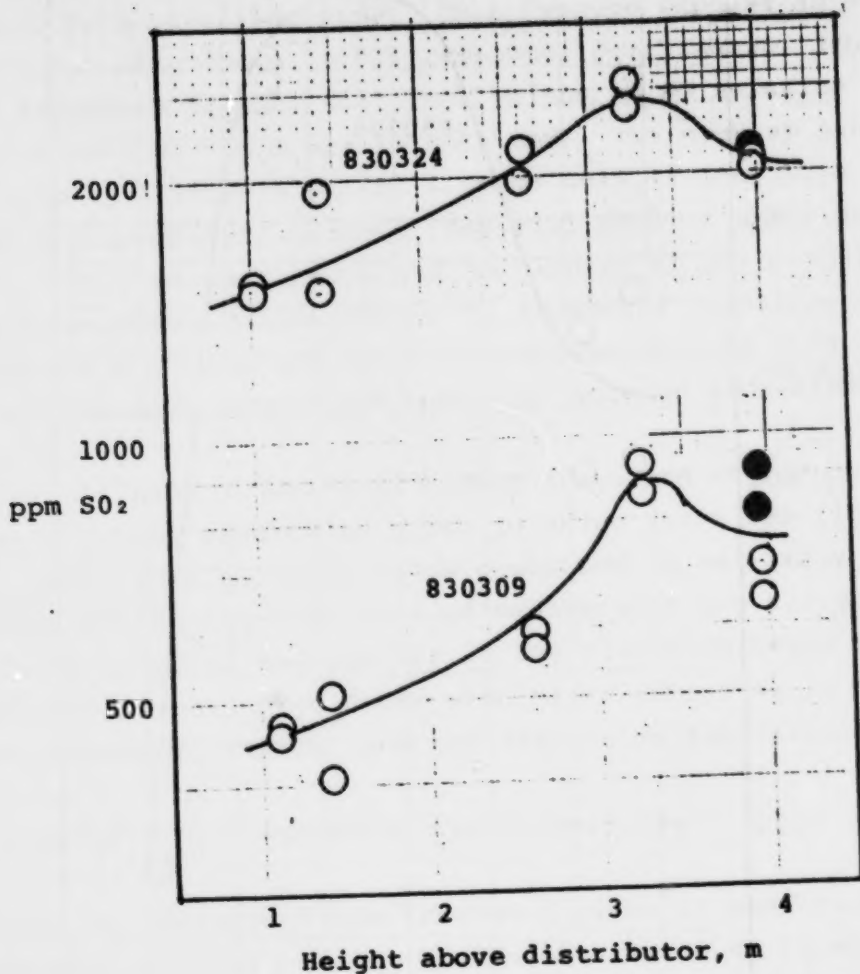


Fig. 6.5. Freeboard profiles of SO₂ content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

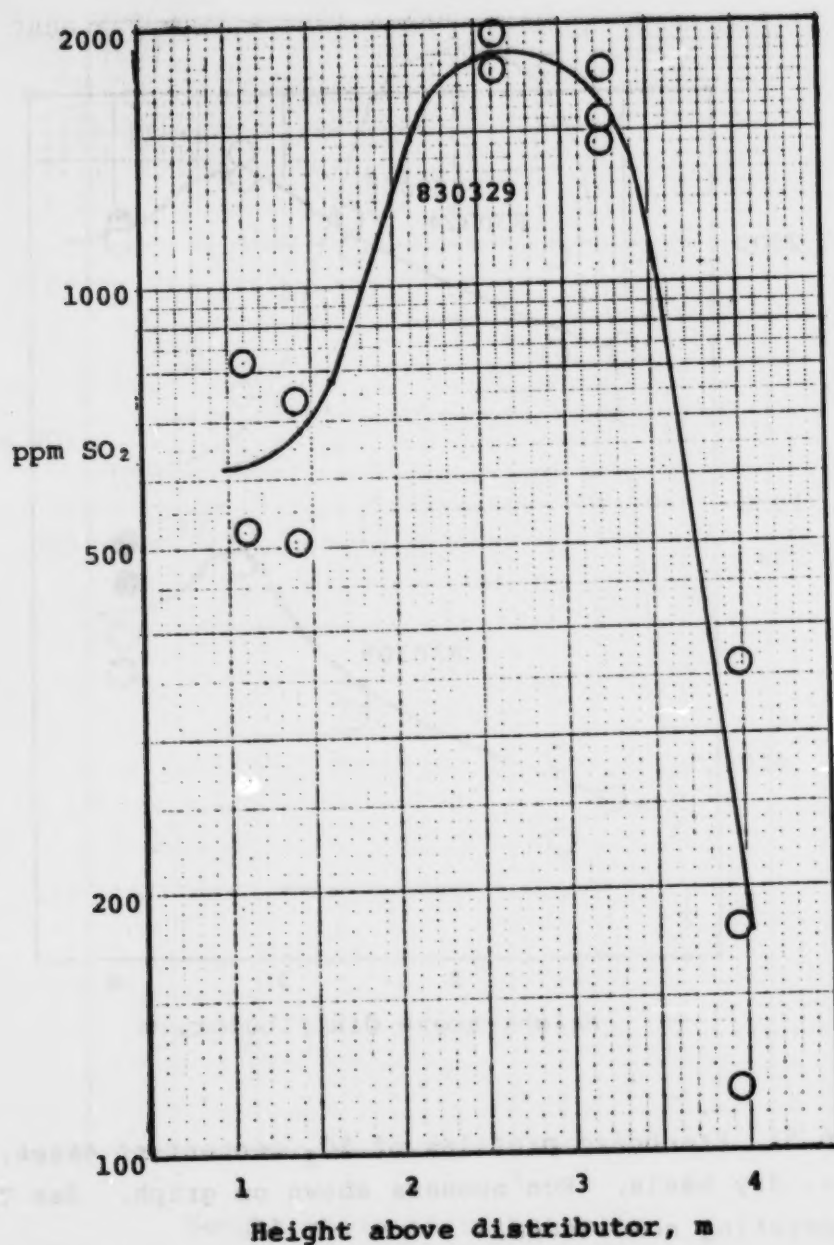


Fig. 6.6. Freeboard profiles of SO_2 content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

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- o freeboard burning of char, producing SO₂
- o freeboard capture of SO₂ by sulphation of CaO
- o freeboard decomposition of CaSO₄, producing SO₂
- o freeboard adsorption or desorption of SO₂ by solid particles, taking from or adding to the SO₂ content of the gases.

- o measurement errors, mainly due, very likely, to failure to quench the reactions quickly enough in the gas sampling probe. However, our design assessment indicates that such errors should be small.

The correct explanation of these results is not simple and will not be pursued here. Two things are abundantly clear, though:

1. Freeboard variation of SO_2 content is substantial
2. This variation introduces another element of apparatus-dependency of system performance.

6.5 Freeboard gas composition profiles: O_2 and CO_2

Figures 6.7 and 6.8 show that the O_2 content of the freeboard gases generally decreased with height. This is the expected effect of freeboard burning, and it is clear that such burning was usually substantial.

As in the case of temporal variation at the combustor exit, the sum of the O_2 and CO_2 contents of the freeboard gas was a constant in a given run, the same in value as at the combustor exit.

6.6 Freeboard gas composition profiles: CO

The behaviour of the CO content of the freeboard gases, Figs. 6.9--6.11, was as complicated as that of SO_2 , and more erratic. Levels as high as 5000 ppm and as low as 200 ppm were observed.

6.7 Closure

The freeboard gas composition profiles demonstrate the considerable importance of freeboard reaction. Rough qualitative

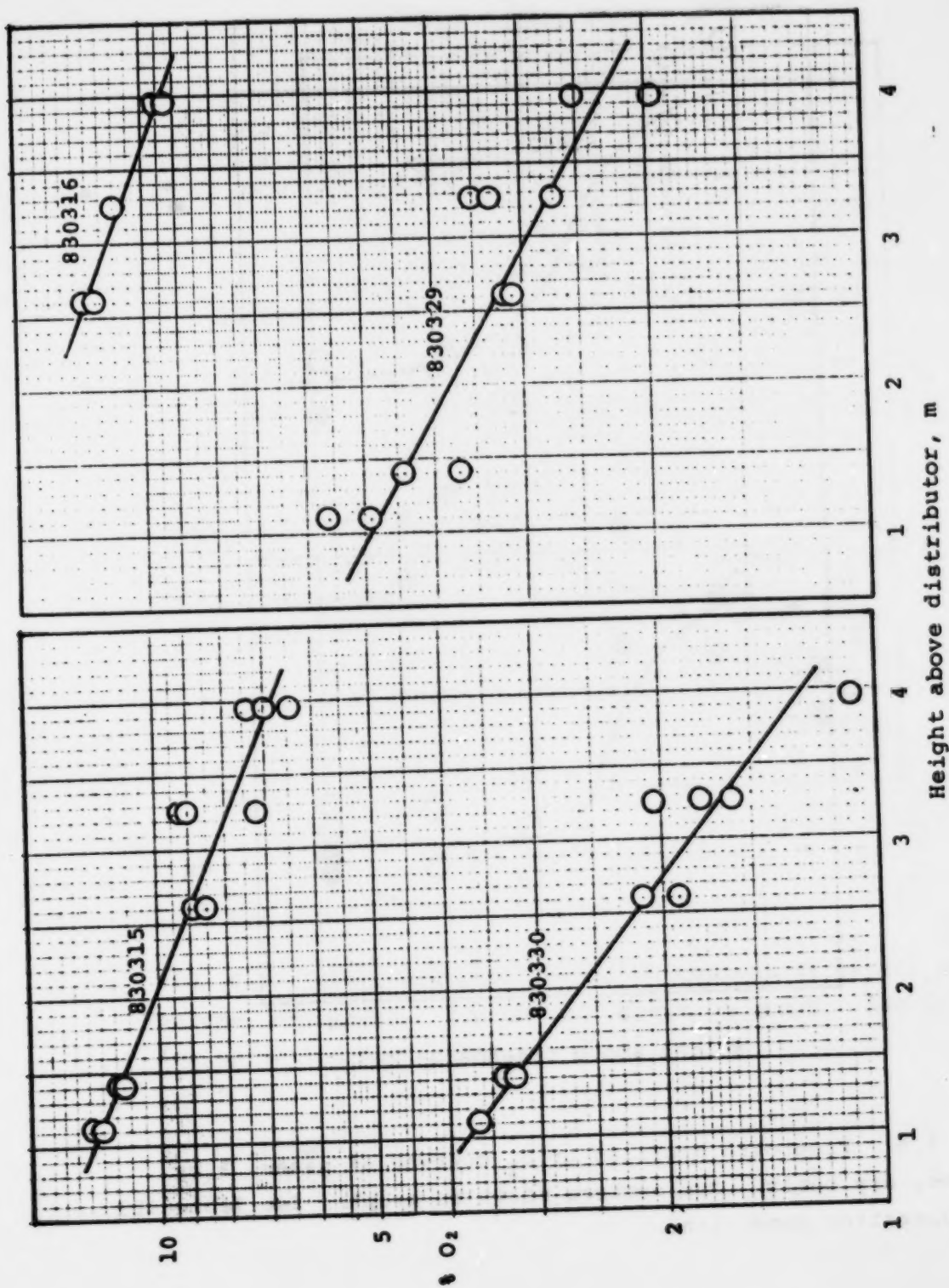


Fig. 6.7. Freeboard profiles of O₂ content of gases, % by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

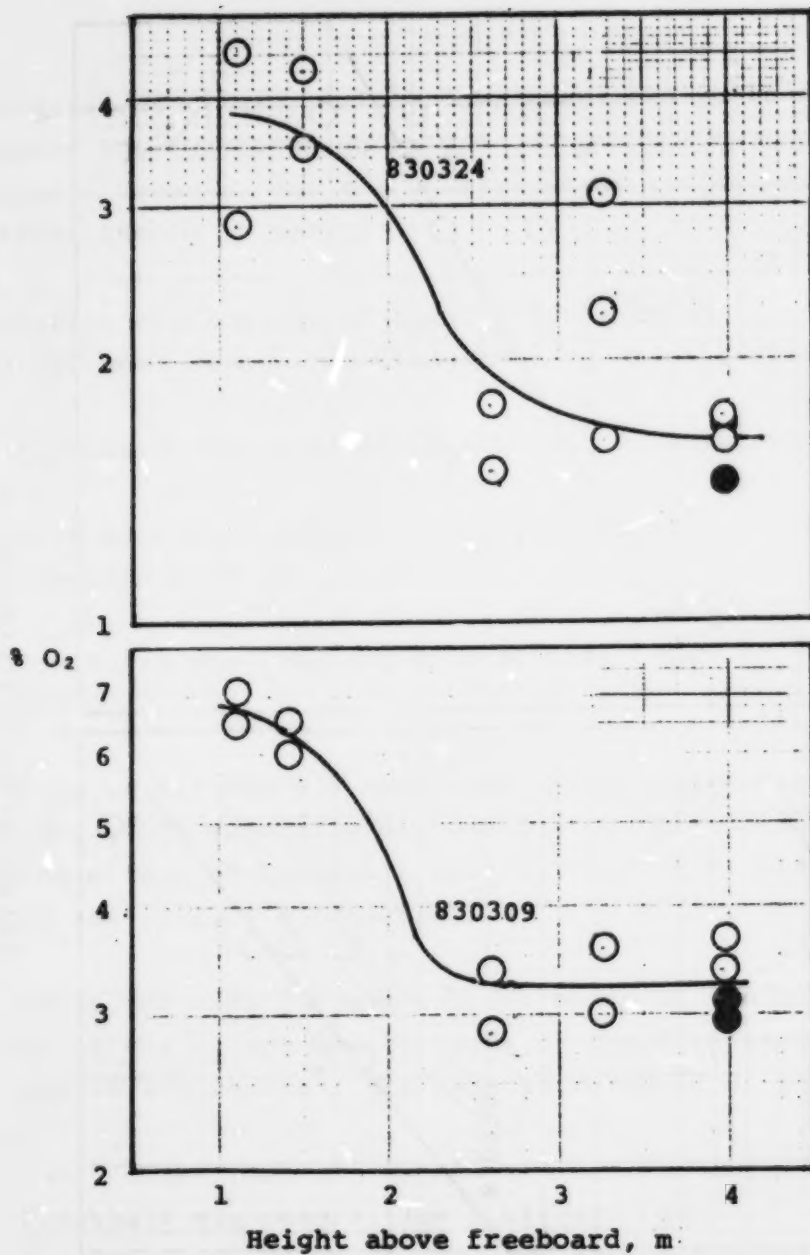


Fig. 6.8. Freeboard profiles of O₂ content of gases, % by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

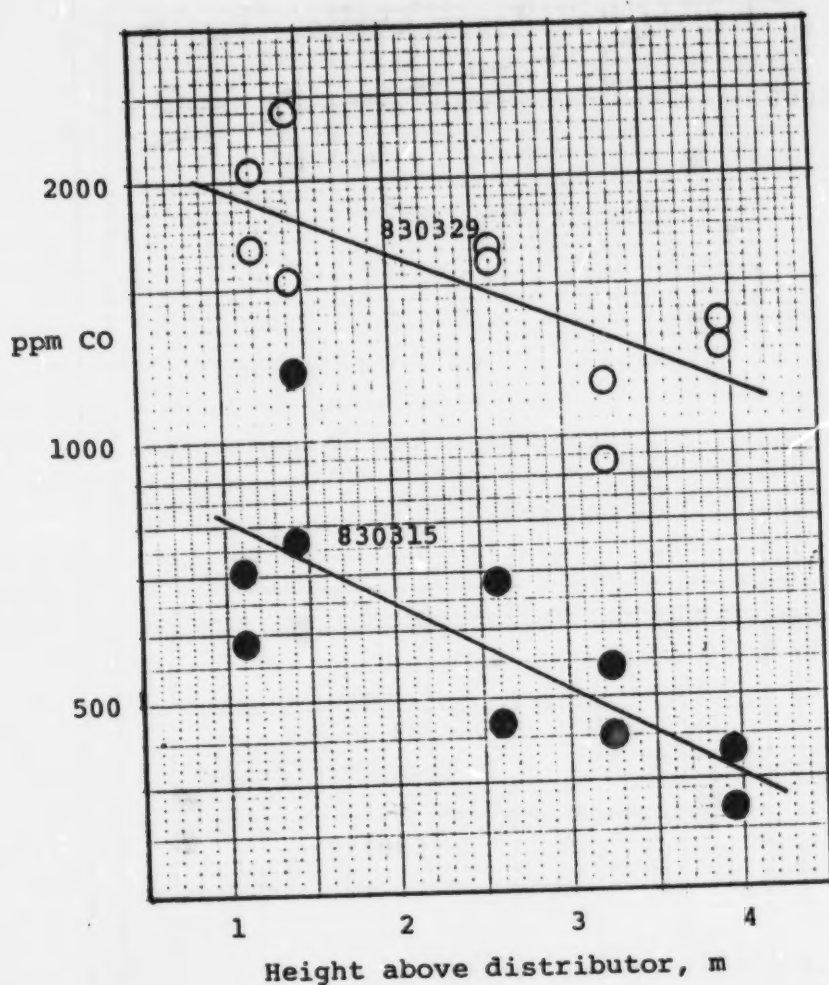


Fig. 6.9. Freeboard profiles of CO content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

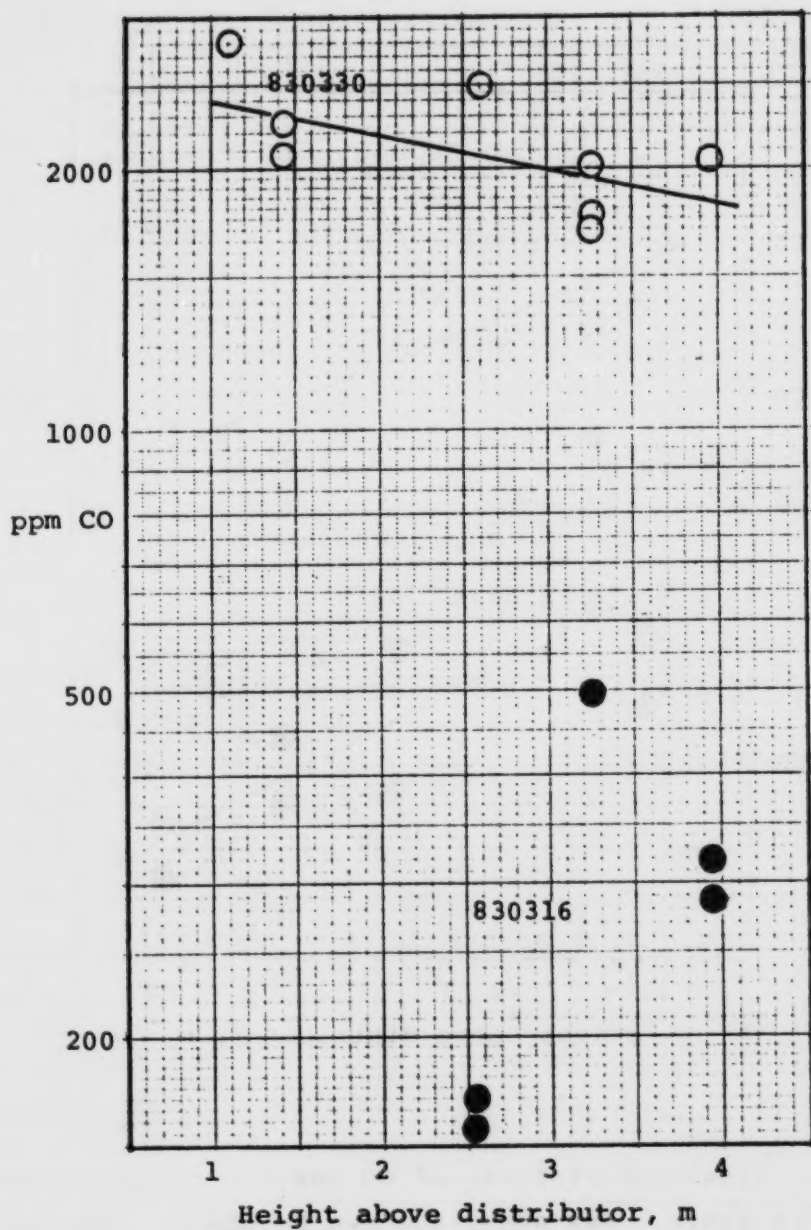


Fig. 6.10. Freeboard profiles of CO content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

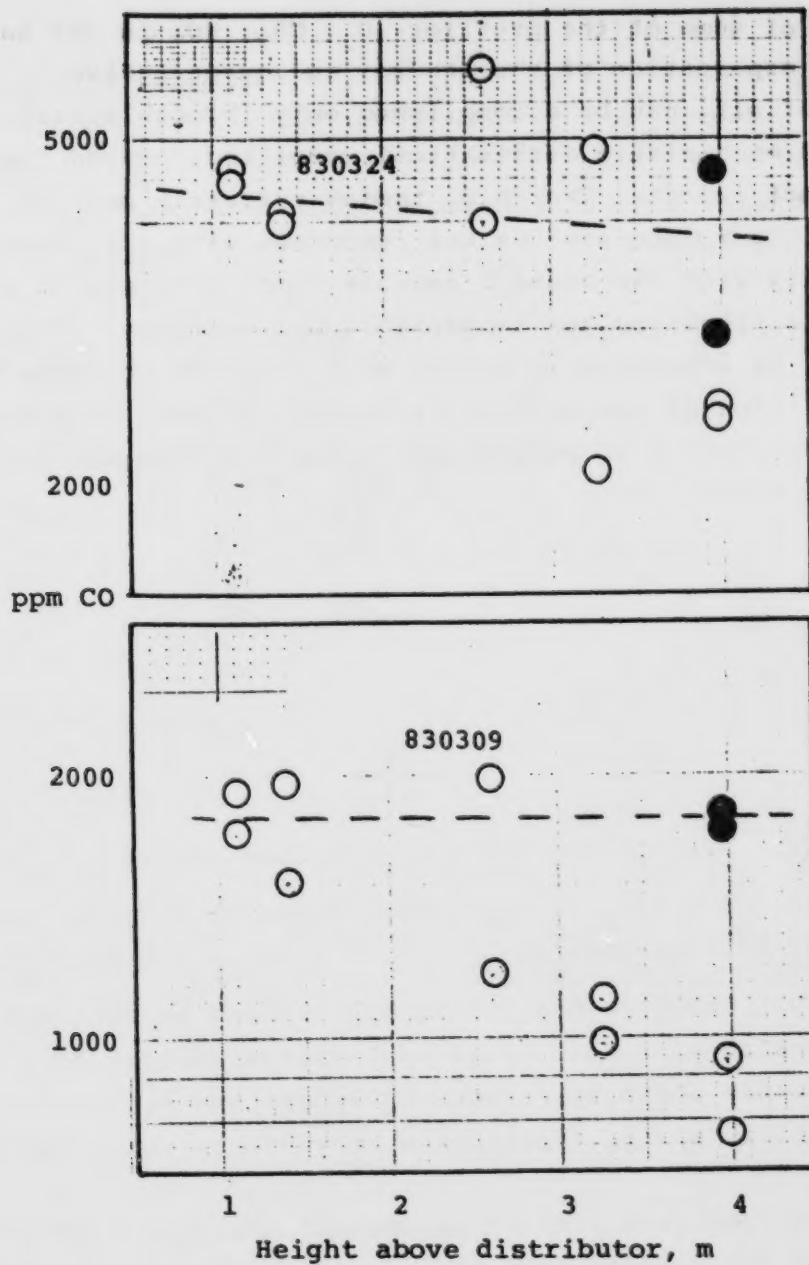


Fig. 6.11. Freeboard profiles of CO content of gases, ppm by volume, dry basis. Run numbers shown on graph. See Table 1.1 for operating conditions.

explanation of some of the profiles (O_2 , CO_2 , NO) is not hard. Qualitative explanation of the others, and quantitative accounting of all, can be accomplished only through appropriately detailed and mechanistic mathematical modelling, beyond the scope of the present report. Two Ph.D. theses currently nearing completion in our laboratories are concerned with such modelling. One difficulty with the present results in this regard is that we lack reliable freeboard gas temperature measurements. This problem will be addressed in future work as noted in Sect. 6.2. With the addition of new vertically lowered probes, reasonably reliable measurements of particulate concentrations and fluxes will also be facilitated.

7. THE SOLIDS BALANCE

7.1 Introduction

A fairly accurate input/output solids flux balance can be written as follows:

$$\begin{aligned} \dot{m}_f w_{\text{ash},f} + \dot{m}_s (1 - w_{\text{CO}_2,s}) \\ = \dot{m}_b (1 - w_{\text{CM},b} - w_{\text{SO}_3,b}) \\ + \dot{m}_c (1 - w_{\text{CM},c} - w_{\text{SO}_3,c}) \\ + \dot{m}_h (1 - w_{\text{CM},h} - w_{\text{SO}_3,h}) , \end{aligned} \quad (7.1.1)$$

where $w_{\text{ash},f}$ is the mass fraction ash in the coal, $w_{\text{CO}_2,s}$ is the mass fraction of bound (carbonate) CO_2 in the limestone, and $w_{\text{CM},n}$ and $w_{\text{SO}_3,n}$ ($n = b, c, h$) are respectively the mass fractions of combustible matter and of bound (sulphate) SO_3 in the n 'th product stream ($n = b, c, h$). This balance was satisfied within the limits of cumulative error statistically indicated by the accuracy of measurement of the various terms.

The main question considered in this Section, however, is the division of the product solids flux between the three product streams: bed product, cyclone product, and baghouse product.

7.2 The recycle system

In the early part of the present work, the recycle control was of the "on-off" type, i.e., either all or none of the cycloned dust was recycled. It soon became evident that this might not allow the best-designed experiments, and indeed

sometimes the cycloned dust flow might become so heavy if none were bled off as product that the recycle system would plug from overload. If the cyclone is highly efficient under given operating conditions, then it can happen under total recycle that very little dust escapes to the baghouse. If at the same time very little elutriable dust is flushed out with the bed product, then the conditions exist for a great elevation of dust hold-up in the system and for a correspondingly heavy burden on the cyclone and the recycle system. Figure 7.1 illustrates such an occurrence where the recycle dust flow rose exponentially to a level of eight times the CaO equivalent of the limestone feedrate, 2.7 times the coal feedrate, and 16 times the coal-ash input rate. We do not know how much further this growth might have gone because a little later a clinker appeared and quickly grew, evidently immobilizing much dust, and this soon forced shut-down. The maximum recycle flux reached was near the limit of the ability of the recycle system to operate without plugging, and indeed the flow could be maintained only by using such a high transport air velocity as to result in unacceptably rapid wear of metal parts at critical points.

A splitter valve was then designed and installed in the dustline under the cyclone, as described in Sect. 2.5, to recycle a fraction r of the cycloned dust and bleed off as product the remaining fraction $1 - r$, $0 \leq r \leq 1$. In the subsequent runs, the recycle dust flux $\dot{m}_r$ ranged up to 50 kg/hr (up to 1--2 times the coal feedrate). The transport air flux employed was 18--27 kg/hr, and the air velocity in the recycle line was 22--32 m/s. We are planning in future work to perform some runs with heavier recycle than is permitted by the present system. A line of nearly double the current capacity will be installed for such experiments. Up to 1/3 of the airflow entering the bed will then come from the recycle port. We are disinclined to supply any greater proportion of the air by this route, since air so served is poorly distributed for fluidization and gas-solids contact.

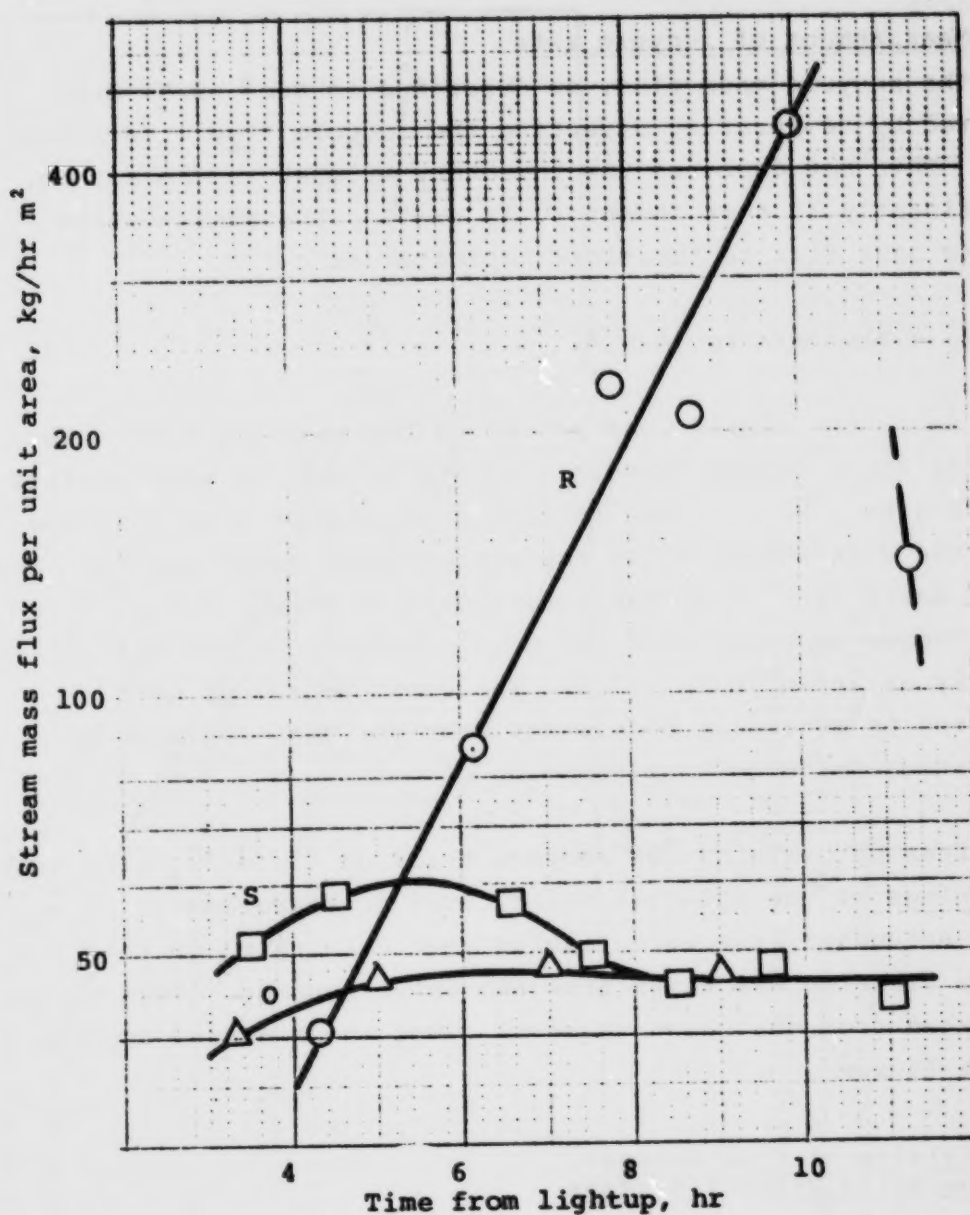


Fig. 7.1. Example of operation with catastrophically growing recycle (run 821207). Run was terminated by the further catastrophe of clinkering at 11 hr. Key to data: O, recycle; □, CaO in limestone fed; Δ, bed product.

7.3 Measurement of recycle rate

The recycle mass flux, as described in Sect. 2.5, was measured by temporarily diverting the recycle flow into a sampling cell. This act depletes the inventory of dust in the system, significantly if that inventory is small. The steady-state recycle mass flux is therefore estimated as

$$\dot{m}_r = \lim(\Delta m / \Delta t), \Delta t \rightarrow 0, \quad (7.3.1)$$

where Δm is the sample mass and Δt is the sampling time. The sampling times ranged from 15 s to 240 s, and the sample sizes from 50 g to 1 kg. It was initially not appreciated just how small the dust inventory in the system might sometimes be. Some of the early data were thus interpreted as shown in Fig. 7.2. It later became apparent that the most plausible interpretation is actually as drawn in Fig. 7.3. The limit behaviour here indicated is generally consistent with our data and with a mathematical model of the sampling process.

After determining the recycle fraction $r = \dot{m}_r / \dot{m}_d$ at a number of settings of the splitter valve, further measurements of $\dot{m}_r$ became redundant, the valve having been calibrated so that the recycle fraction was known from the valve setting. Then $\dot{m}_r$ can be calculated from the known value of r and the measured cyclone product flux $\dot{m}_c$.

7.4 Division of flow between the solid product streams

Figure 7.4 shows, for all runs in which steady state operation was obtained, the division of solid product between the three product streams: bed product, cyclone product, and baghouse product. Except in a few runs with total recycle of cycloned dust, the baghouse product never exceeded 13 % of the total, and fell as low as 3 %. The cyclone product composed 11--57 % of the total, and the bed product 39--85 %. In the one fully successful run at total recycle, the baghouse product made up 57 % of the total.

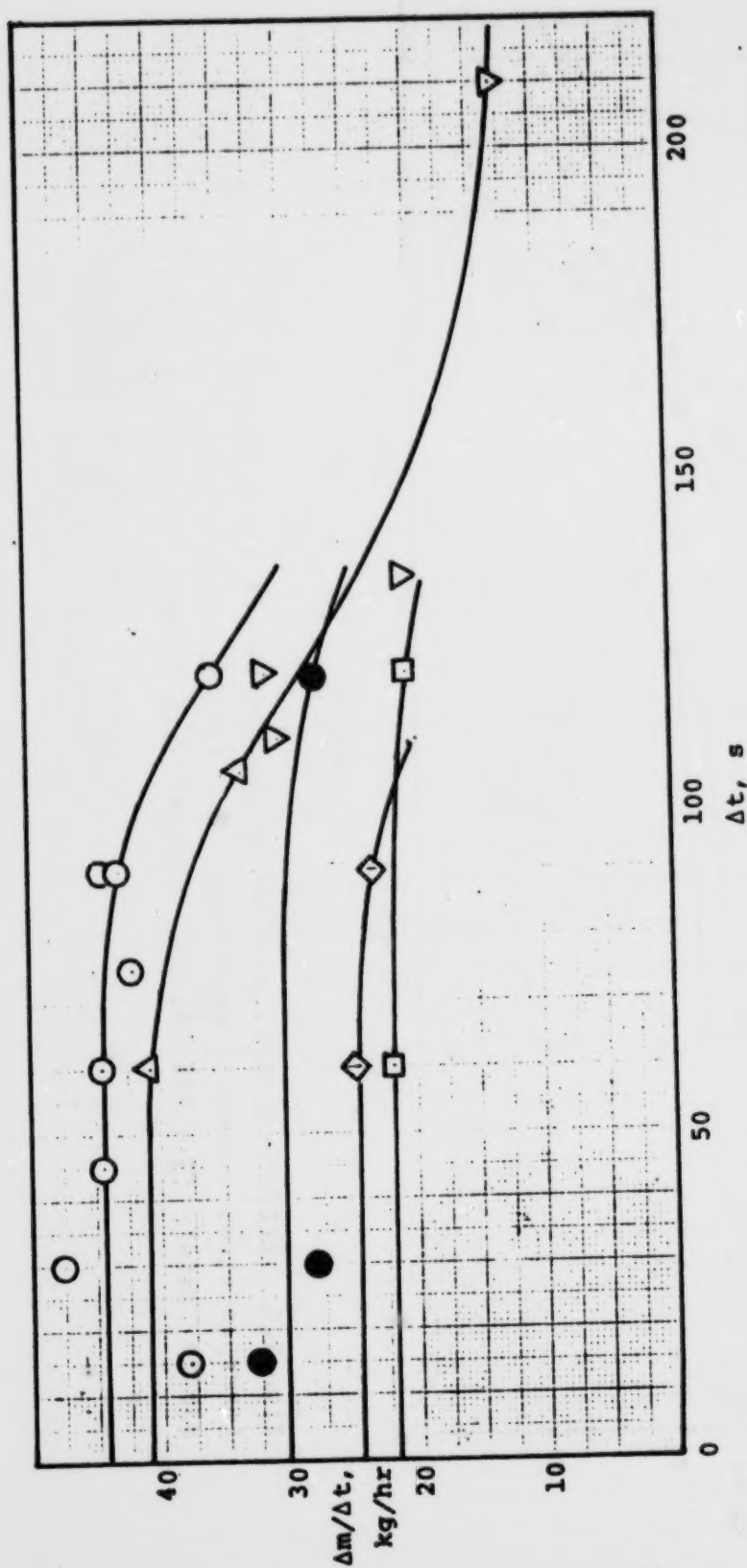


Fig. 7.2. Effect of sampling time on recycle sample size, first interpretation. Runs in question: $\bullet$, 830324 (early); $\circ$, 830324 (late); $\diamond$, 830224; $\triangle$, 830202; ∇ , 830112.

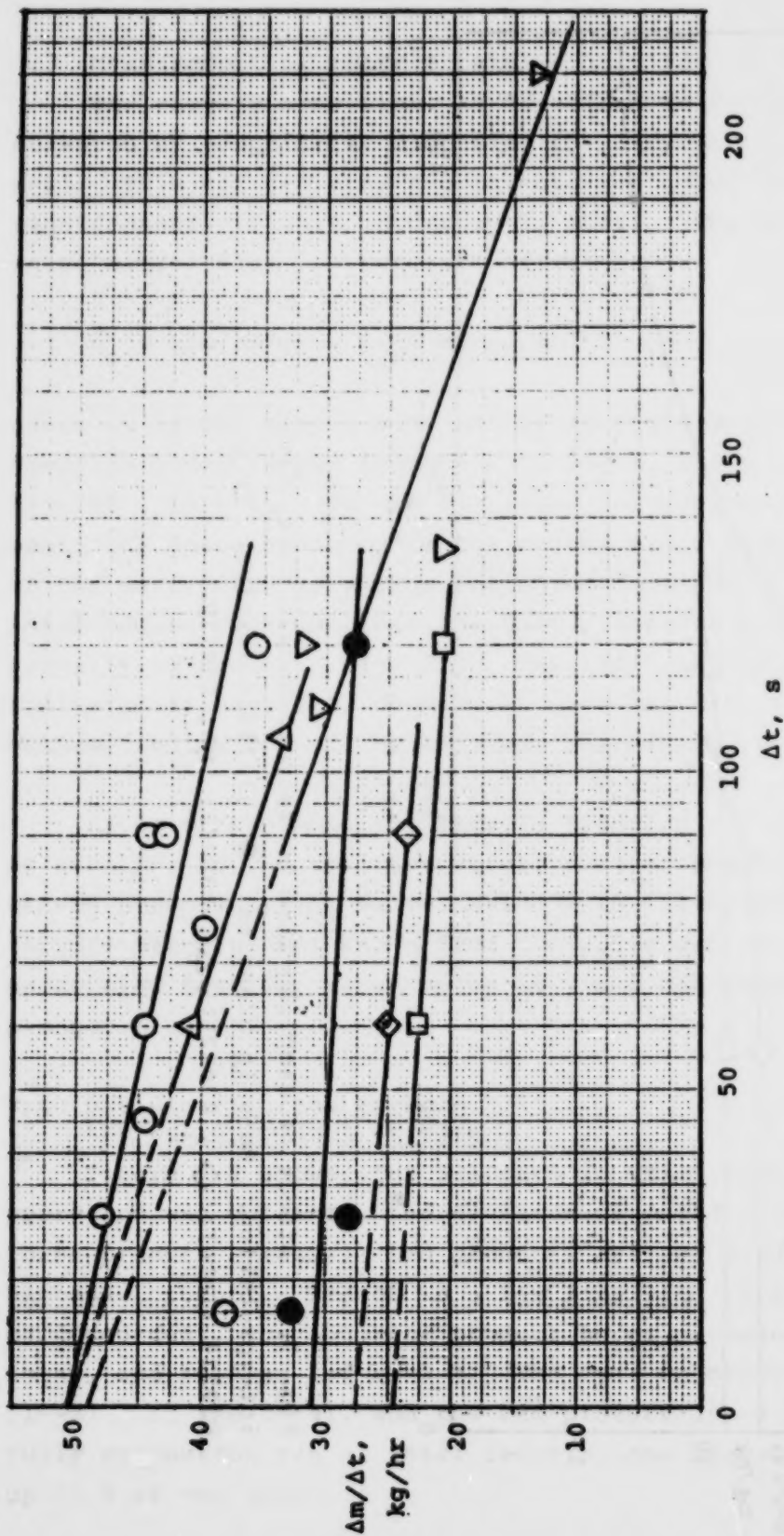


Fig. 7.3. Effect of sampling time on recycle sample size, second interpretation. Runs in question: ●, 830324 (early); ○, 830224 (late); □, 830310; ◇, 830224; △, 830202; ▽, 830112.

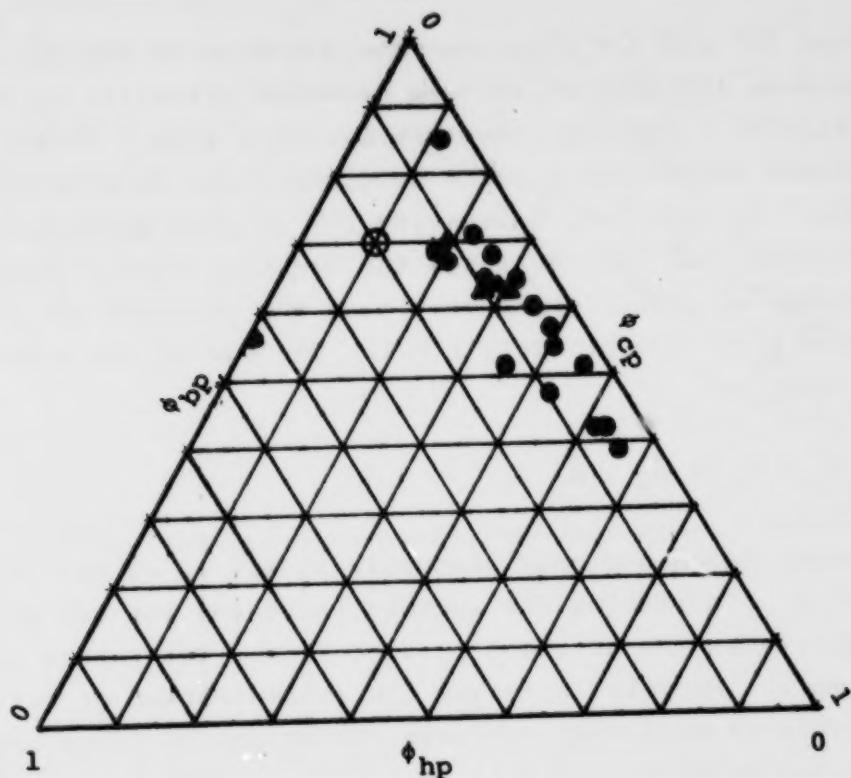


Fig. 7.4. Partition of solid products between bed product, cyclone product, and baghouse product, where $\phi_{bp} = \dot{m}_b / \dot{m}_p$, $\phi_{cp} = \dot{m}_c / \dot{m}_p$, and $\phi_{hp} = \dot{m}_h / \dot{m}_p$. Circled point (illustrates reading of graph): $\phi_{bp} = 0.7$, $\phi_{cp} = 0.1$, $\phi_{hp} = 0.2$. Data points: $\bullet$, Minto trials; $\blacktriangle$, Devco trials.

Figures 7.5 and 7.6 show that the division of the solid product between the product streams depended primarily on the recycle fraction r and the limestone particle size. These figures indeed constitute a quite good empirical correlation of the results. As expected, increasing the recycle decreases the cyclone product flux and increases the baghouse flux. These changes become extremely sharp very near the point of total recycle of the cyclone product, $r = 1$. The sum of the overhead product fluxes,

$$\dot{m}_o = \dot{m}_c + \dot{m}_h = \dot{m}_p - \dot{m}_b ,$$

however, seems to change only gradually at all r , as may be seen from Fig. 7.6. Increasing the recycle increases the bed product flux and decreases the overhead product flux. The reason for this is simple. Recycle increases the concentration of fines in the bed. This in turn must increase the amount of fines carried out in the bed product.

The effect of increasing limestone particle size, as expected if the particle size distributions are of similar form, is to increase the bed product and to decrease both the cyclone product and the baghouse product.

A simple explanation of the observed behaviour is as follows. The solids leaving the system originate in the limestone and the coal ash, apart from a modest amount of char. We might expect that in operation with a given ash content, the division of the solid products would depend largely on the particle size distribution of the limestone, the feed ratio of coal to stone, the fluidizing velocity, and the recycle ratio. The results thus suggest that the influences of coal/stone ratio and the fluidizing velocity are so modest that they are here imperceptible.

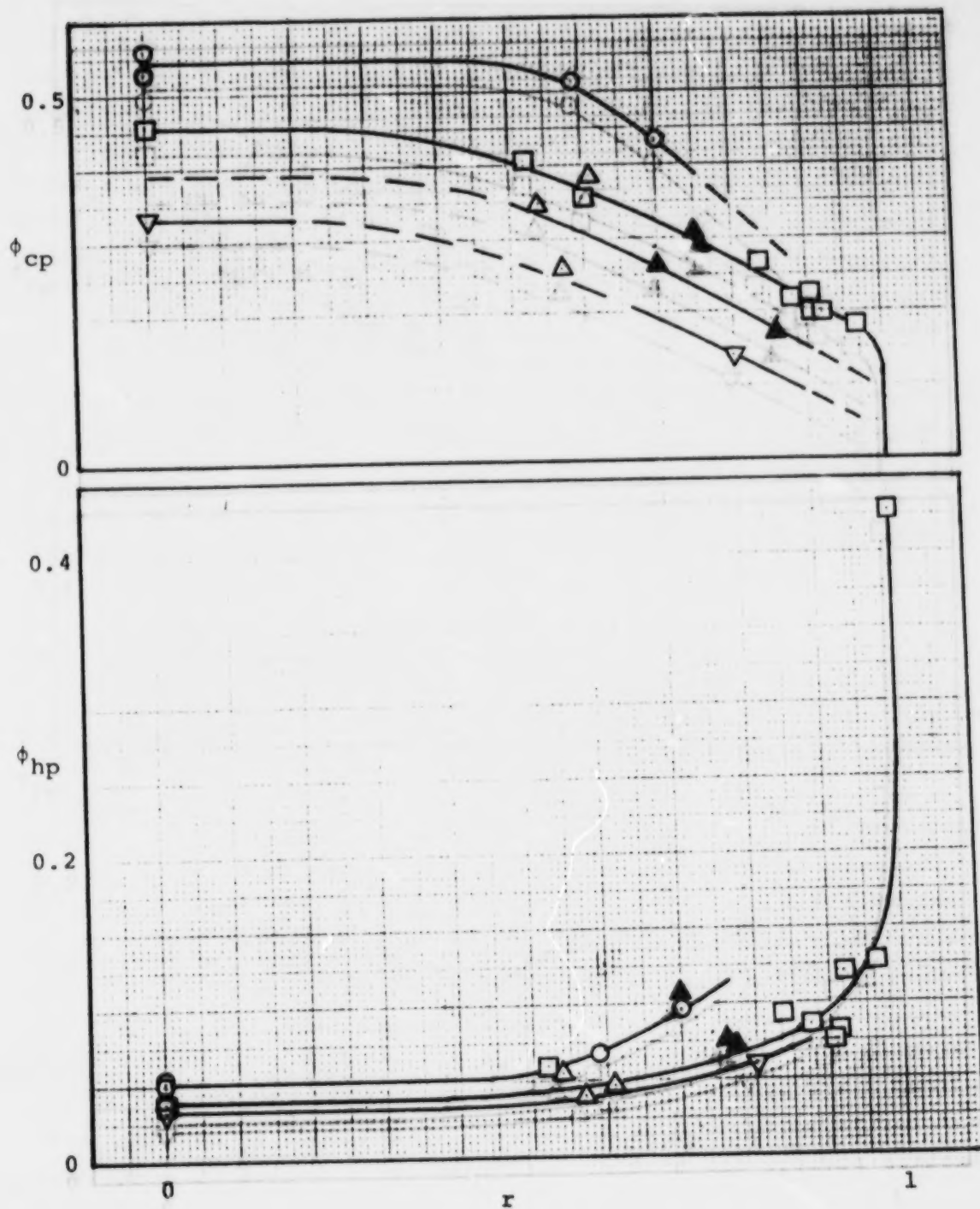


Fig. 7.5. Flux ratios of cyclone (top) and baghouse (bottom) products to total solids product, vs. recycle fraction. Key to data: Table N.3A.

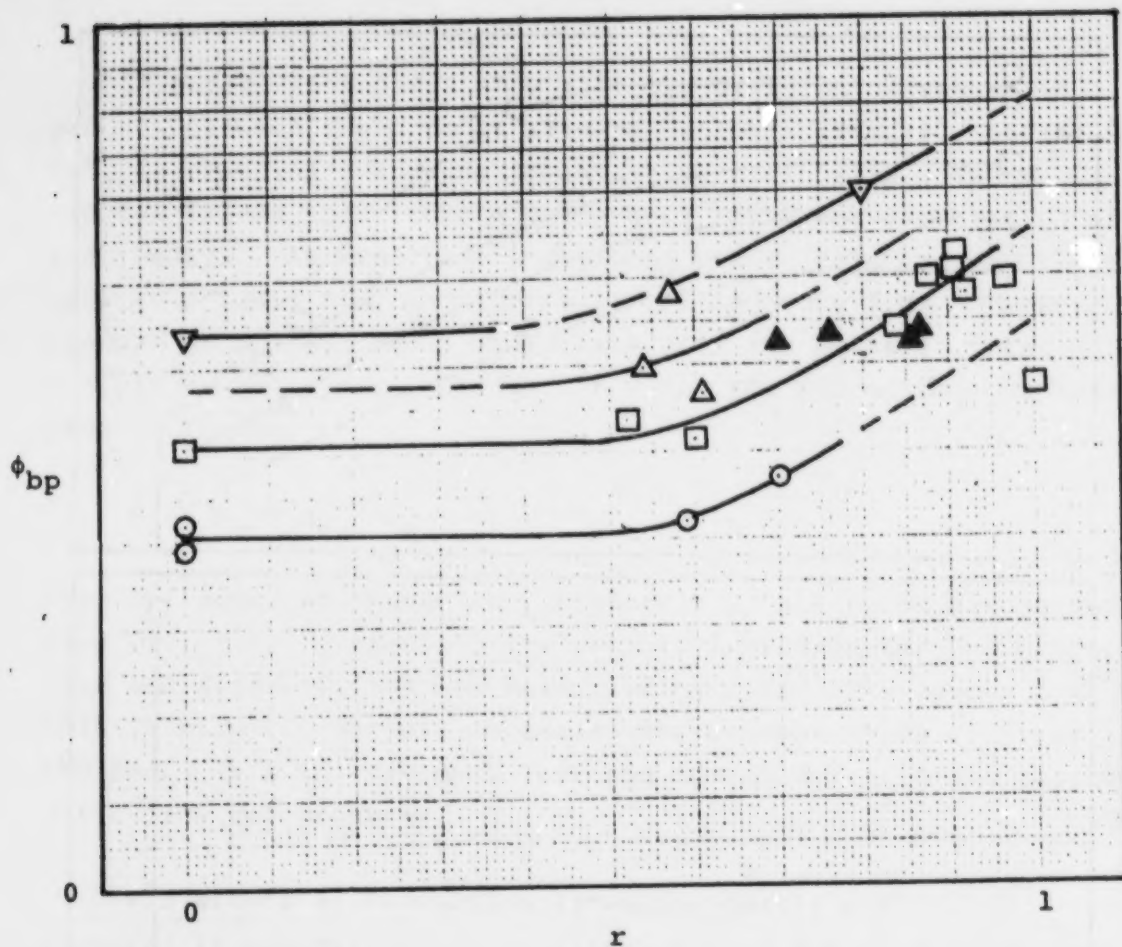


Fig. 7.6. Flux ratio of bed product to total solids product, vs. recycle fraction. Key to data: Table N.3A.

The division of the overhead product between cyclone product and baghouse product in the neighborhood of $r = 1$ is not quantitatively conveyed by Figs. 7.5 and 7.6. It is, however, well indicated by Fig. 7.7 which shows a strong correlation between the flux ratio $\dot{m}_c/\dot{m}_h$ and the recycle fraction. The equation of the line statistically fitted to the data for $\hat{D}_p = 0.49$ mm and 0.78 mm is

$$\dot{m}_c = 11 \dot{m}_h (1 - r)^{0.68} \quad (7.4.1)$$

It is clear that the division of the overhead solid products depends on the cyclone performance and is thus apparatus-dependent. The theory of cyclone separation of dusts can be invoked to predict this performance. The pursuit of such an inquiry is, however, beyond the present scope.

7.5 Division of flow resolved with respect to particle size

Figure 7.8 shows for one representative run the particle size distributions of the solids fed (coal and limestone) and the solid products withdrawn (bed product, cyclone product, baghouse product, and the combination of these three). The data on product size distribution together with those on the product mass fluxes (Sect. 7.4) can be interpreted to give the distribution of product mass flux with particle size. Consider the flux ratio $\phi_{np} F_{D,n} = \dot{m}_n F_{D,n} / \dot{m}_p$, where $\dot{m}_n F_{D,n}(D_p)$ is the mass flux of that part of the n 'th solids stream that consists of particles of diameter D_p less than the argument value. An example illustrating the distribution of $\phi_{np} F_{D,n}$ with particle diameter D_p for the various product streams is shown in Fig. 7.9. Another very interesting flux ratio, denoted by $\phi_{D,np}$, characterizes the size-resolved partition of mass flux between the product streams. We define

$$\phi_{D,m} = \dot{m}_m \frac{\partial F_D}{\partial D_p} \quad (7.5.1)$$

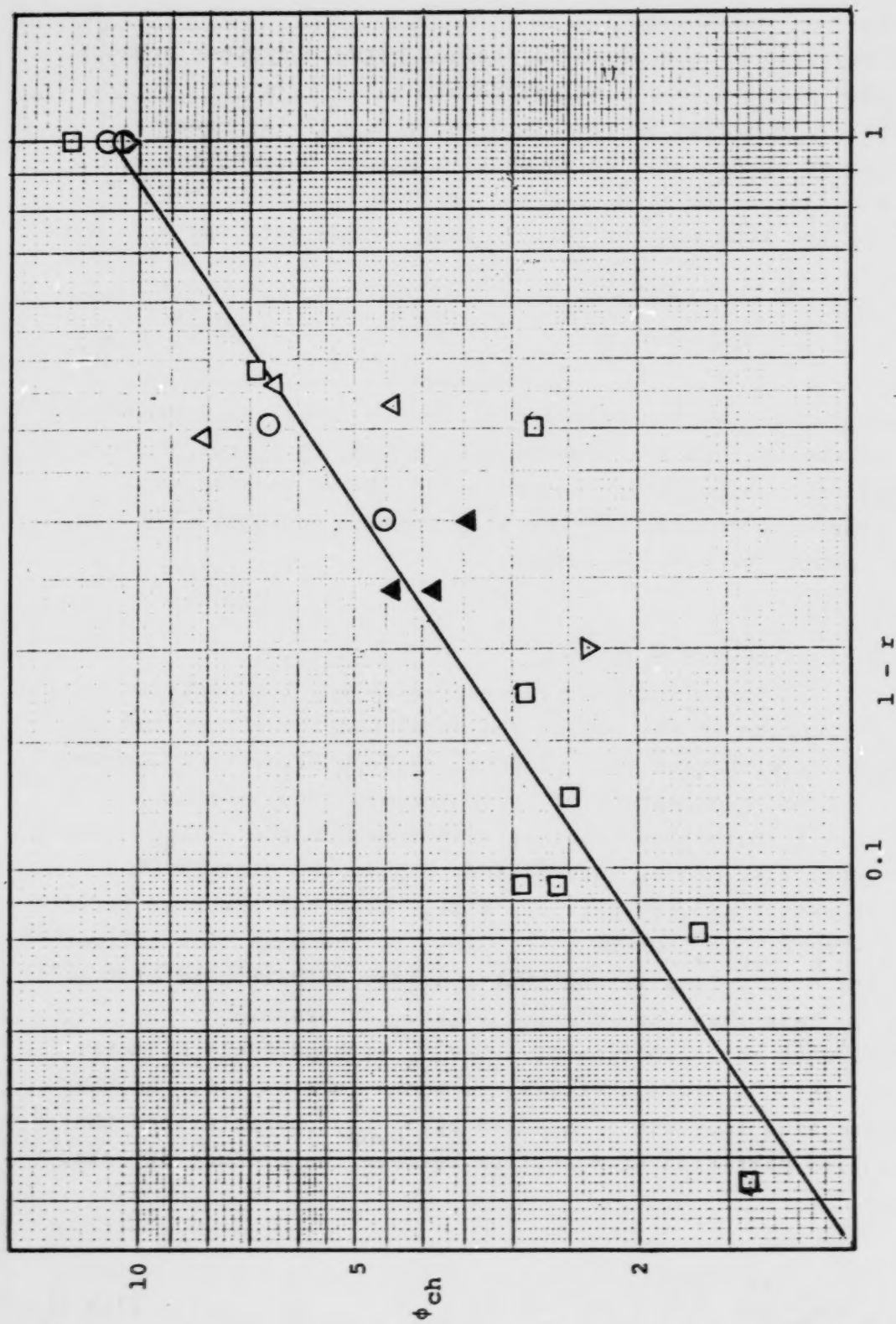


Fig. 7.7. Flux ratio of cyclone product to baghouse product, vs. recycle fraction. Key to data: Table N.3A.

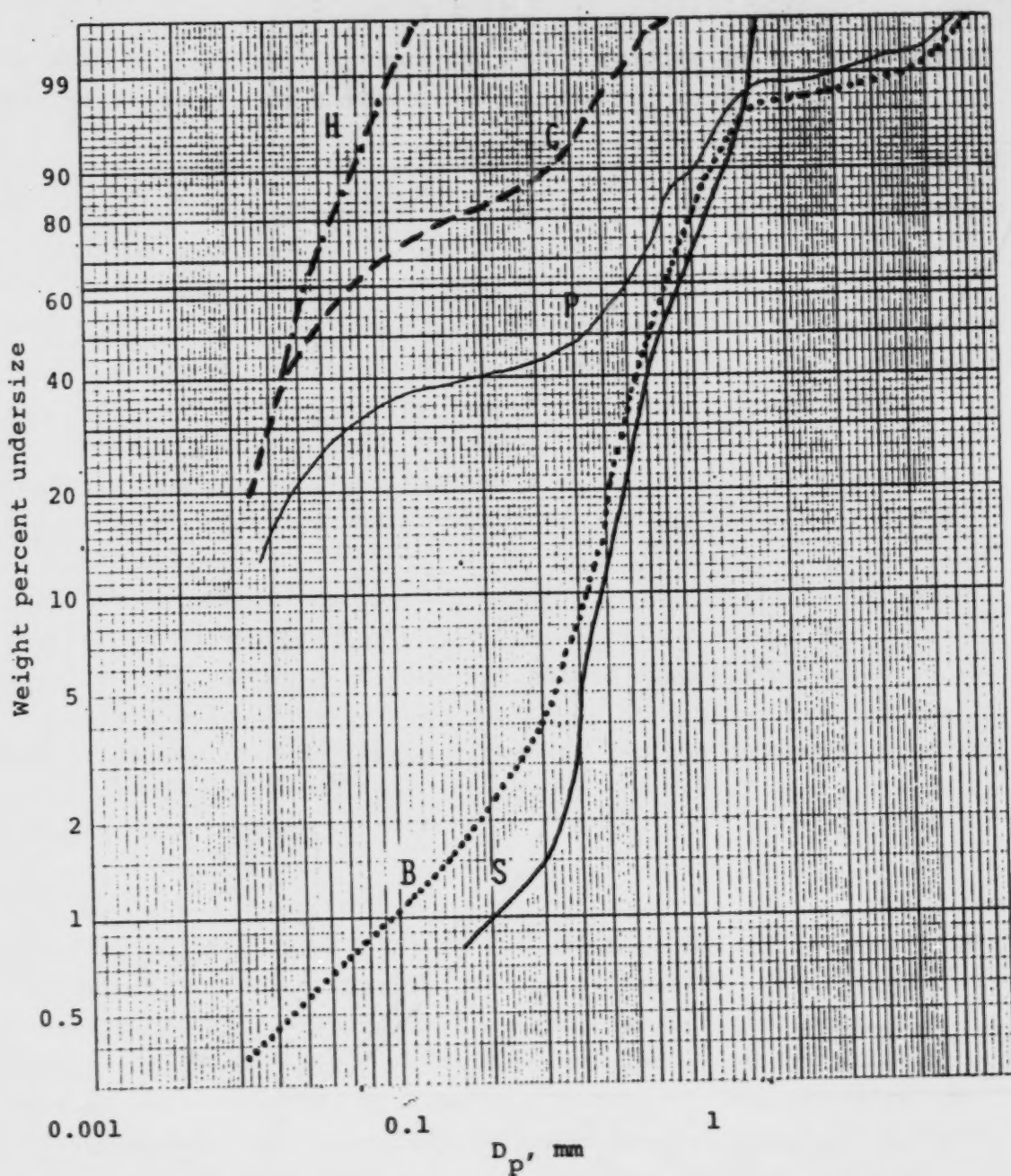


Fig. 7.8. Particle size distributions of solids, run 830216.
Curves: B, bed product; C, cyclone product; H, baghouse product; P, total product; S, limestone.

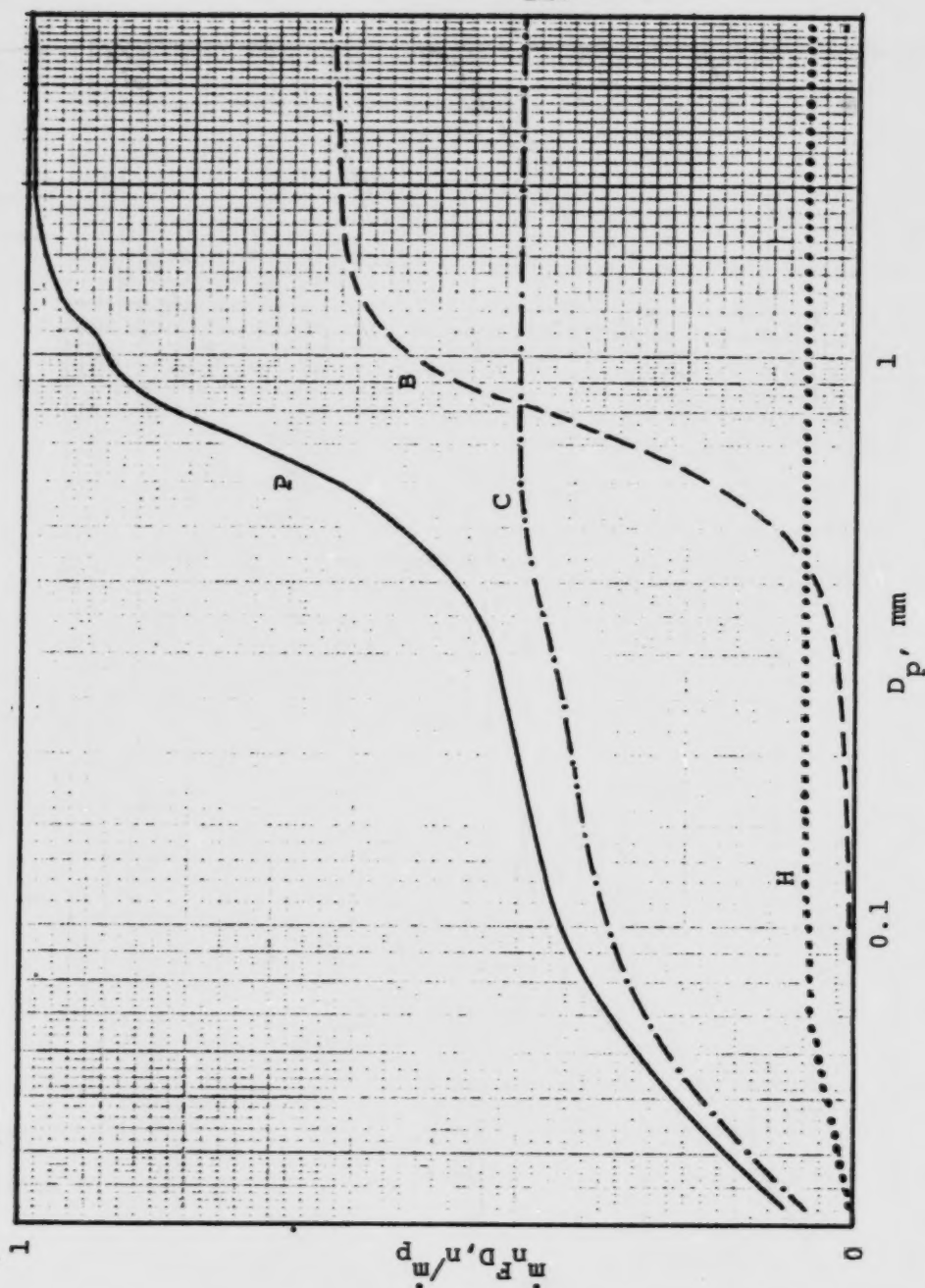


Fig. 7.9. The flux distribution function $\dot{m}_n F_{D,n} / \dot{m}_p$ vs. particle diameter D_p , where $F_{D,n}(D_p)$ is the mass fraction of the material of the n 'th stream ($n = b, c, h, p$) consisting of particles smaller than the argument value of D_p . Curves: B, bed product; C, cyclone product; H, baghouse product; P, total product. Data from run 830216. Example: at $D_p = 0.1 \text{ mm}$, 35 % of the total product flux is in particles smaller than 0.1 mm, 29 % is in cyclone product under 0.1 mm, 5 % is in baghouse product under 0.1 mm, and a negligible fraction is in bed product under 0.1 mm.

so that $\dot{m}_{D,m} dD_p$ is the mass flux of particles in the m'th stream having diameters in the range $(D_p, D_p + dD_p)$ and

$$\dot{m}_m = \int_0^\infty \dot{m}_{D,m} dD_p . \quad (7.5.2)$$

Then $\phi_{D,np} = \dot{m}_{D,n} / \dot{m}_{D,p}$ is flux ratio of particles of diameter D_p in the n'th and p (total product) streams. Fig. 7.10 illustrates the behaviour of $\phi_{D,np}$. The results in Figs. 7.9 and 7.10 demonstrate a fairly sharp separation between bed product and baghouse product, but substantial overlap between the baghouse and cyclone products, and between the cyclone and bed products.

The nature of the overlap between product streams is, of course, more complex than Figs. 7.8--7.10 can show. For example, the larger particles of cyclone dust are expected to be of lower average density than particles of similar size in the bed product. The smaller particles of cyclone dust are expected to be of higher average density than particles of similar size in the baghouse product.

7.6 Closure

The material presented in this section basically has to do with elutriation. A Visiting Scientist from the People's Republic of China, Professor B.-Y. Xu, has analyzed most of our results in the manner of Sect. 7.5, and more. He has moreover carried out cold-model studies of elutriation complementing the pilot plant work. These results and a general analysis in terms of a theory of elutriation are presented in a companion report (Xu, Becker and Code 1983).

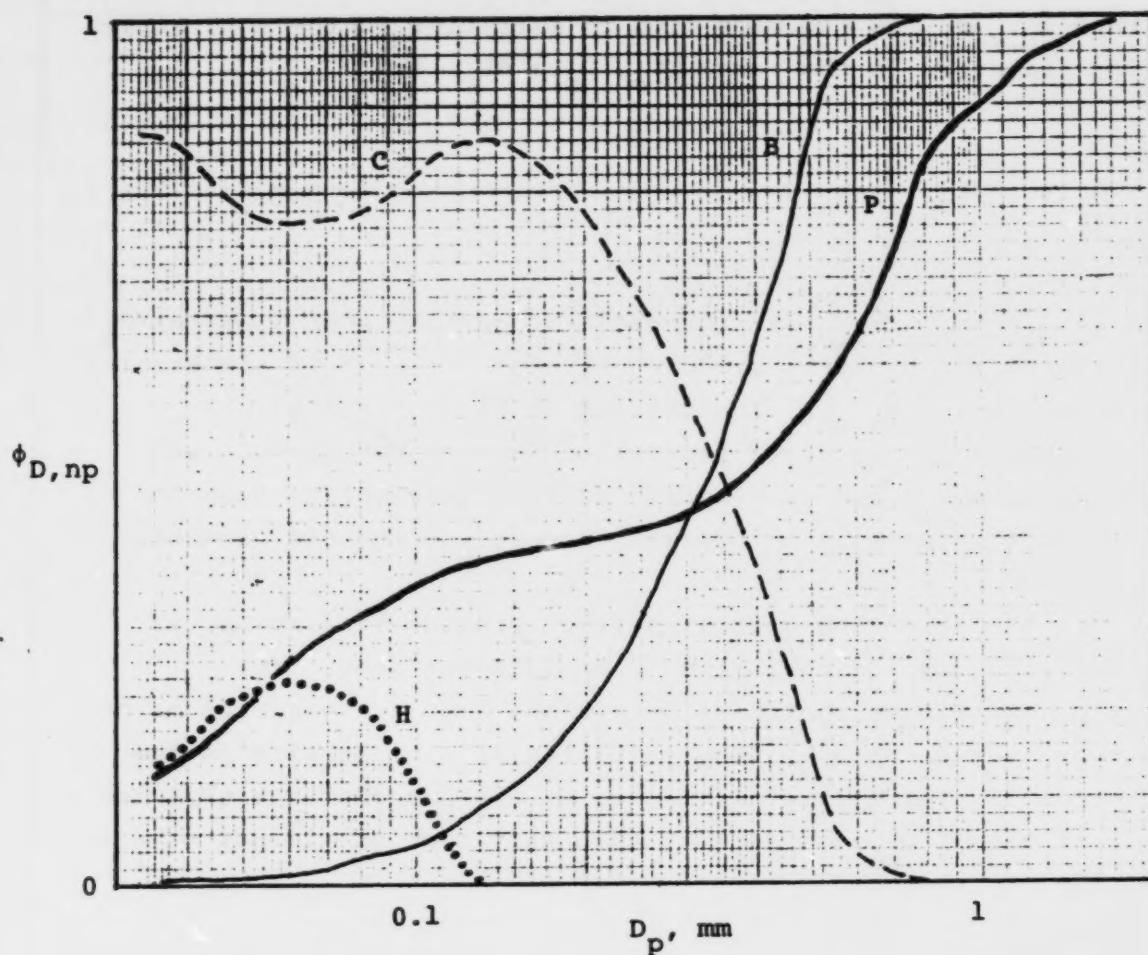


Fig. 7.10. Particle-size-resolved partition of mass flux between solid product streams, run 830216. Curves: B, bed product; C, cyclone product; H, baghouse product; P, total product.

8. THE CARBON BALANCE

8.1 Carbon content of solid products

Figures 8.1--8.3 show the carbon content of the solid product streams as functions of bed temperature, excess air level, and recycle fraction. The ranges of carbon content are: 0.4--3 % for the bed product, 6--30 % for the cyclone product, and 6--18 % for the baghouse product. Clearly, the elutriated material is very high in carbon content relative to the bed. There appear to be tendencies for the carbon content to drop with increasing bed temperature and increasing excess air, as expected if other factors were constant. At the higher recycle levels, $r > 0.6$, there is an apparent tendency for carbon content of the bed and baghouse products to increase with increasing recycle fraction. The significance of these apparent effects, however, cannot be gauged in isolation. Before forming any conclusions, the results must be examined in terms of the carbon fluxes involved. This is done in the following section.

8.2 Carbon fluxes

8.2.1 General remarks

The mass fluxes of carbon carried by the bed product, the cyclone product and the baghouse product are $\dot{m}_{C,b}$, $\dot{m}_{C,c}$ and $\dot{m}_{C,h}$. The total carbon flux carried by the solid product streams is the sum $\dot{m}_{C,p} = \dot{m}_{C,b} + \dot{m}_{C,c} + \dot{m}_{C,h}$. Figure 8.4 accordingly shows the division of the total carbon flux between the three streams. It is seen that the loss of bed carbon with the bed product was generally relatively minor, 3--23 % of the total. The cyclone product was usually the major carrier, taking 59--93 % of the total in nearly all runs. The marked exception, of course, occurred when--at total recycle--no cyclone product was drawn; then the baghouse product took around 80 % of the carbon.

An obviously significant normalization of the carbon fluxes

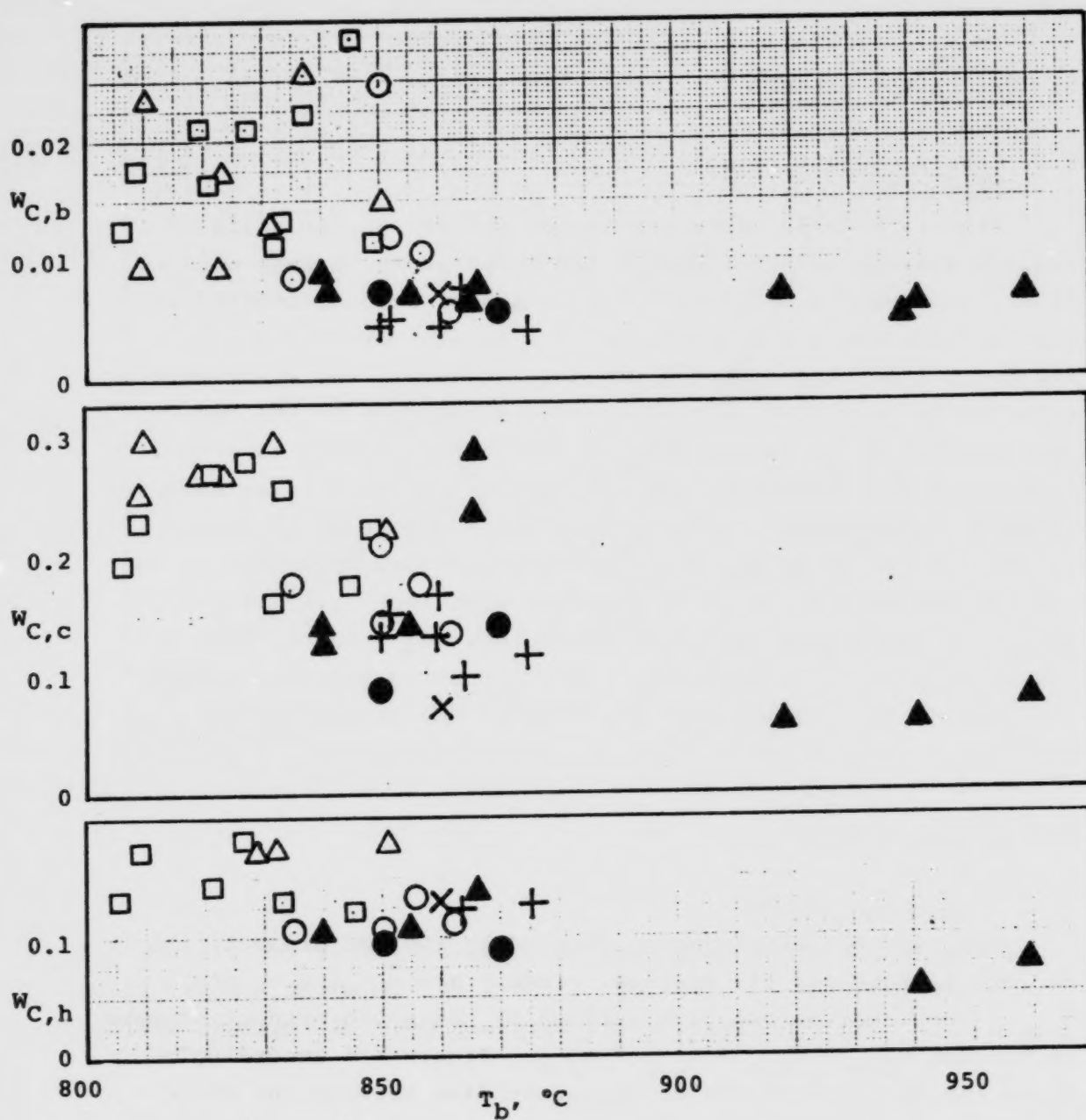


Fig. 8.1. Mass fraction carbon in the solid products vs. bed temperature. Key to data: Table N.3B.

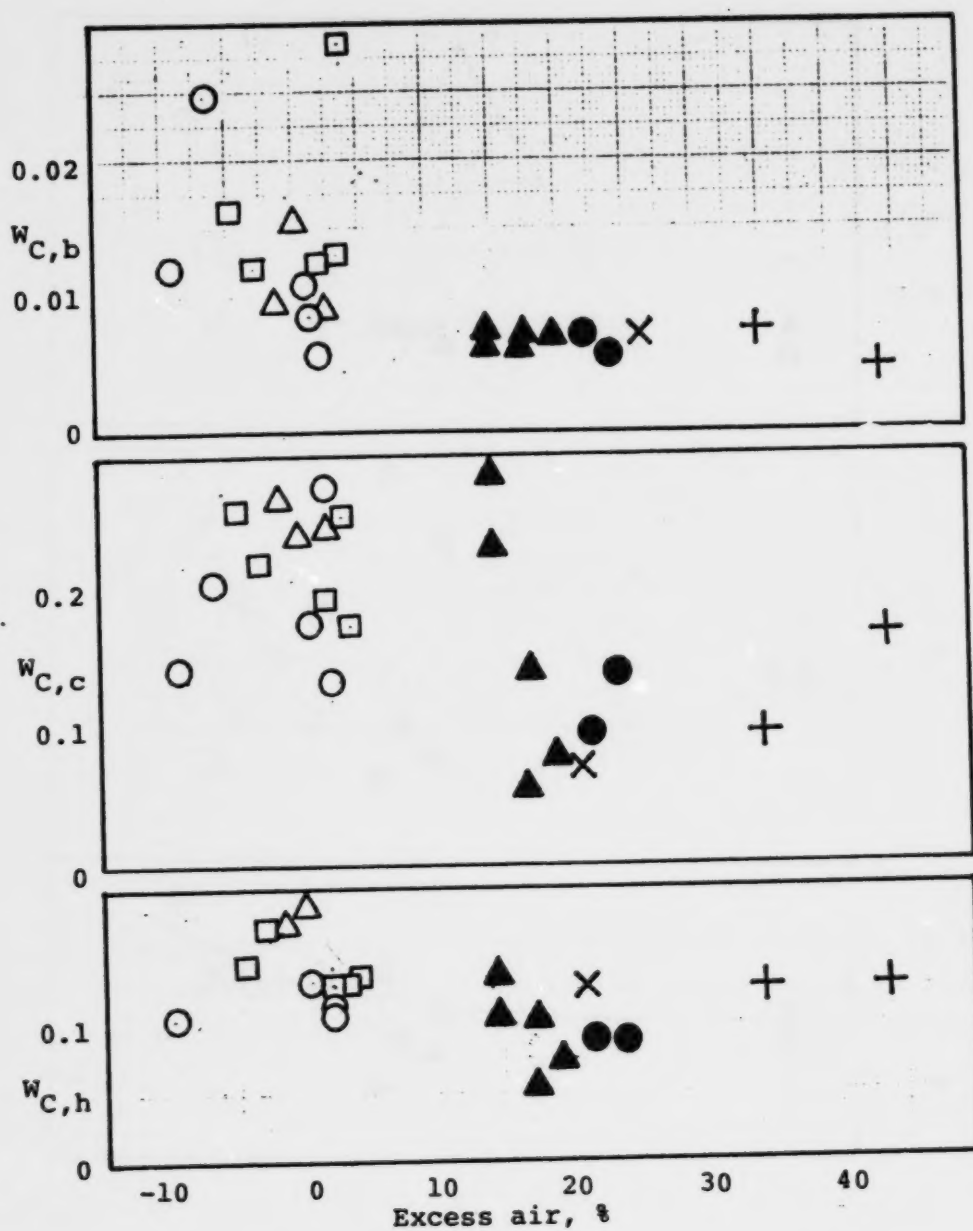


Fig. 8.2. Mass fraction carbon in the solid products vs. excess air level. Key to data: Table N.3B.

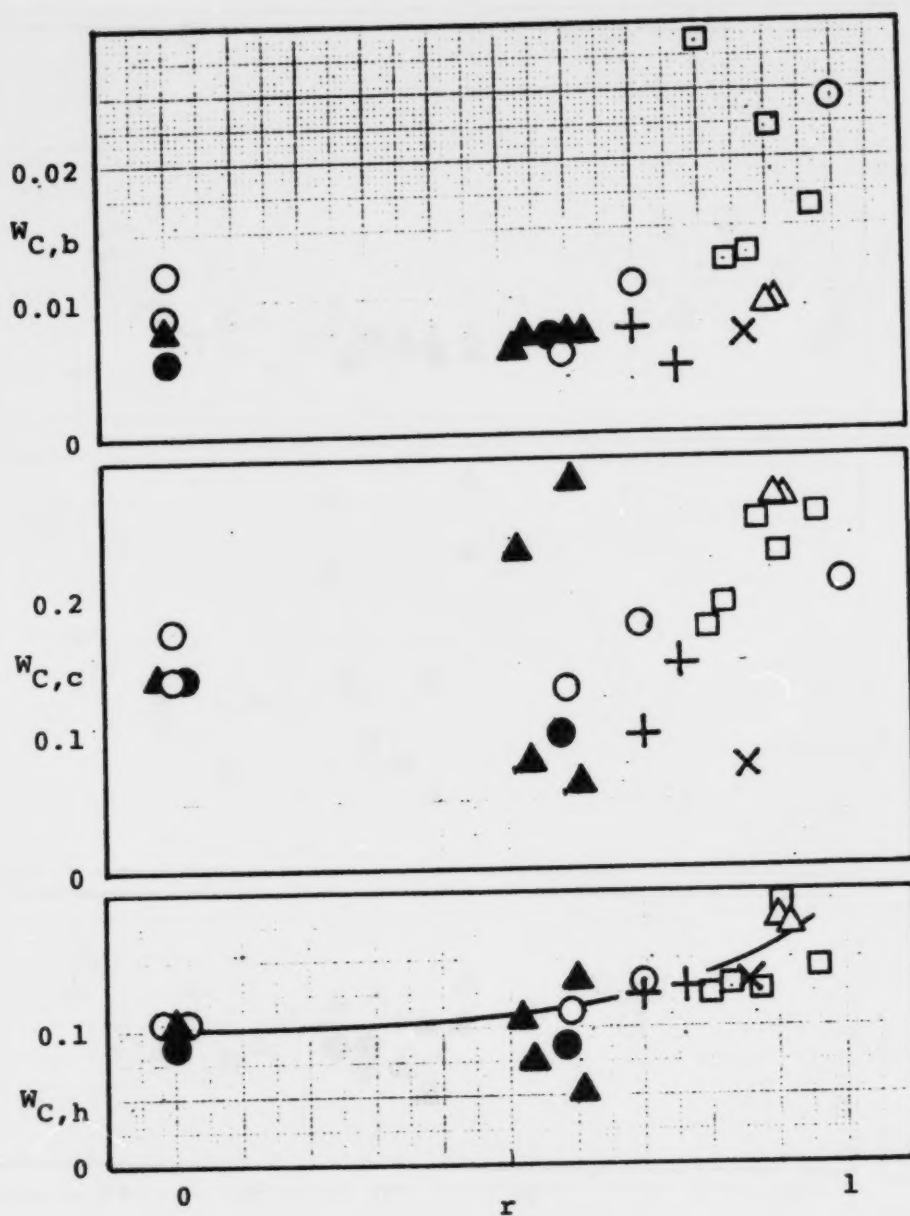


Fig. 8.3. Mass fraction carbon in the solid products vs. recycle fraction. Key to data: Table N.3B.

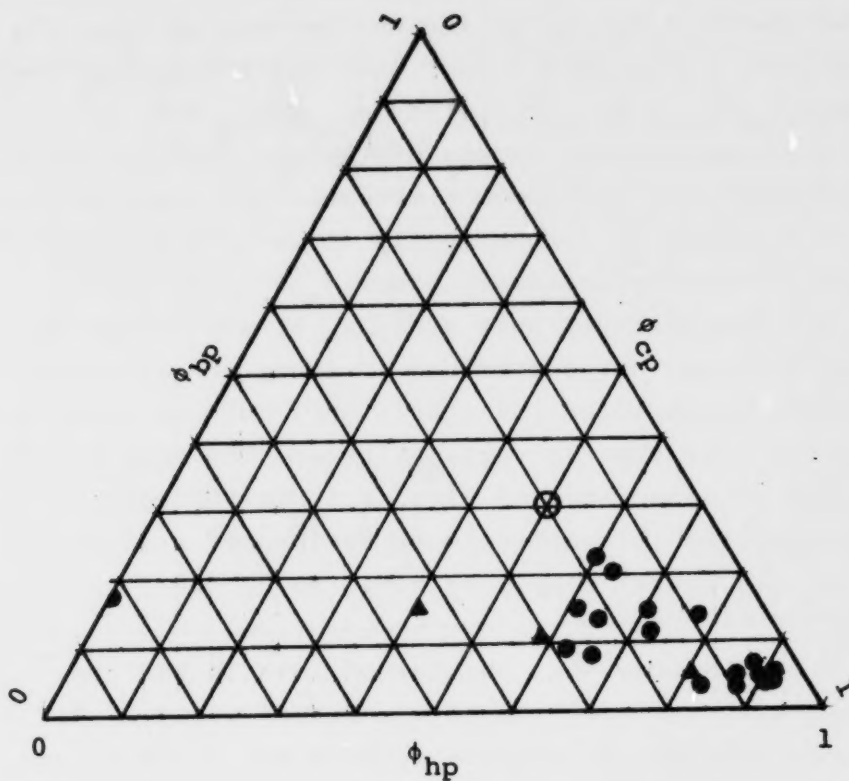


Fig. 8.4. Partition of carbon loss between bed product, cyclone product, and baghouse product. Coals: ●, Minto; ▲, Devco. Illustrative point (circle): $\phi_{C,bp} = 0.7$, $\phi_{C,cp} = 0.5$, $\phi_{C,hp} = 0.2$.

is on the fuel carbon flux, $\dot{m}_{C,f}$, the carbon carried into the system by the coal. Figures 8.5--8.9 show the fractional fuel carbon losses $\dot{m}_{C,b}/\dot{m}_{C,f}$, $\dot{m}_{C,c}/\dot{m}_{C,f}$, and $\dot{m}_{C,h}/\dot{m}_{C,f}$ as functions of bed temperature, excess air level, recycle fraction, total solids efflux, and bed product efflux. The last two actually serve as measures (conveniently in reciprocal form) for the characteristic solids residence times $\tau_p = \dot{m}_b/\dot{m}_p$ and $\tau_b = \dot{m}_b/\dot{m}_b$. The bed mass per unit area, $\dot{m}_b/A$, was typically about 500 kg/m²; it was not determined in every run (because measurement required shutdown for over 6 hr), but the data on bed pressure drop indicate that it varied little. The mass fluxes per unit area on the abscissas of Figs. 8.8 and 8.9 can therefore be converted into approximate reciprocal residence times τ_p and τ_b (in hours) simply by dividing by 500.

Figures 8.5--8.9 show that the carbon loss in the bed product in runs with cyclone product withdrawal was 0.2--1.2 % of the fuel carbon, that in the cyclone product was 0--10 %, and that in the baghouse product was 0.3--3 %. In a run with total recycle of cyclone dust, the loss in the baghouse product rose to 4 %. In general, loss in the bed product was relatively minor. The major avenue of loss was usually through the cyclone product, but at very high recycle levels the loss in the baghouse product became dominant.

Interpretation of the results simply by inspection of Figs. 8.5--8.9 should be approached cautiously, and any conclusions must be regarded as somewhat tentative. There are at least six possibly influential factors at work, namely bed temperature, excess air level, recycle ratio, residence time, coal particle size, and coal type. In addition, fluidizing velocity and limestone particle size may have some effect. In Sects. 8.2.2--8.2.9, we therefore try to resolve the data into sets, such that in each set only one of the apparently strongly influential variables is subject to wide variation. Ultimately the surest method of interpreting the data is with the help of

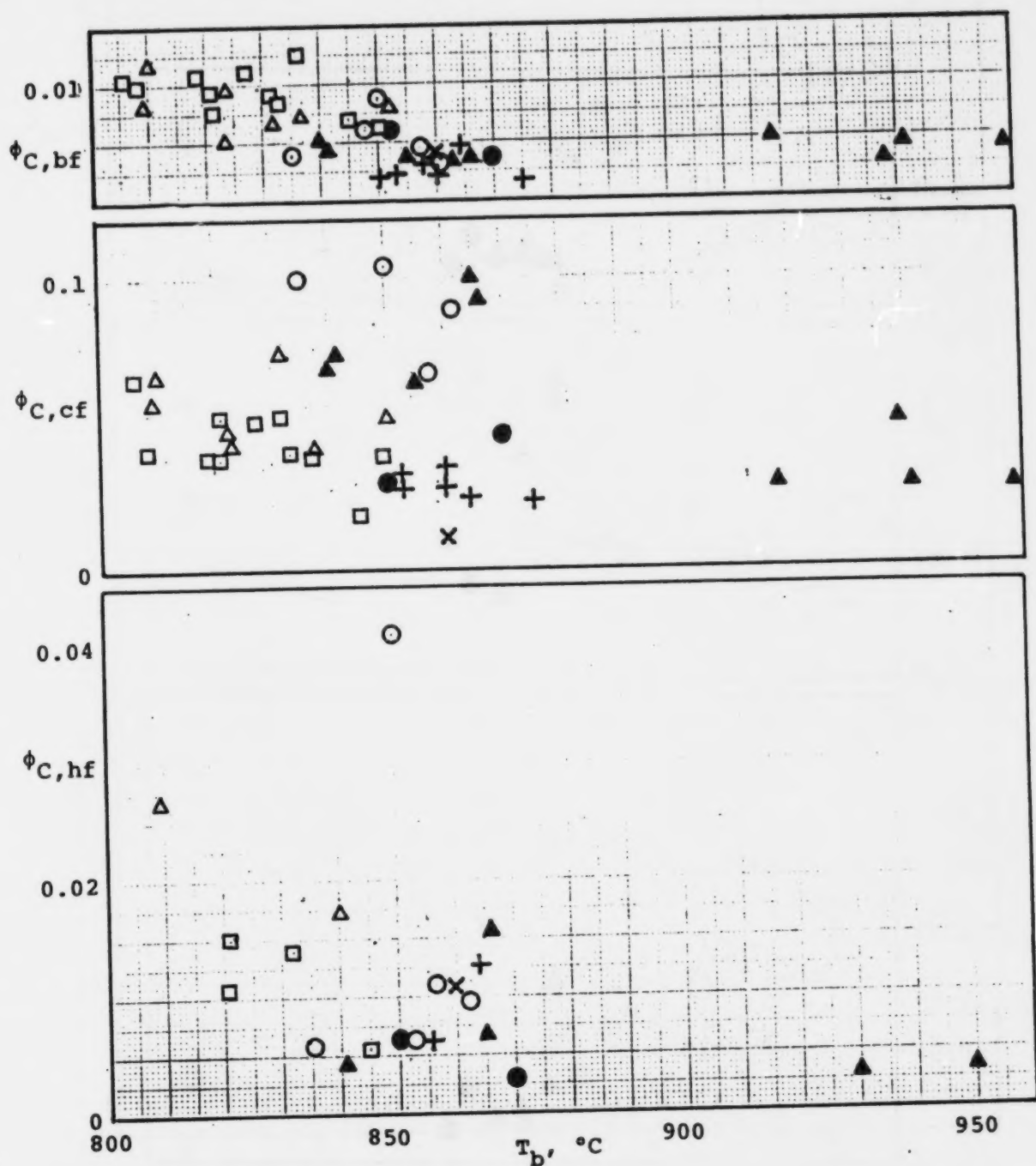


Fig. 8.5. Flux ratio of carbon loss in a product stream to carbon fed in fuel, vs. bed temperature. Top, bed product; middle, cyclone product; bottom, baghouse product. Key to data: Table N.3B.

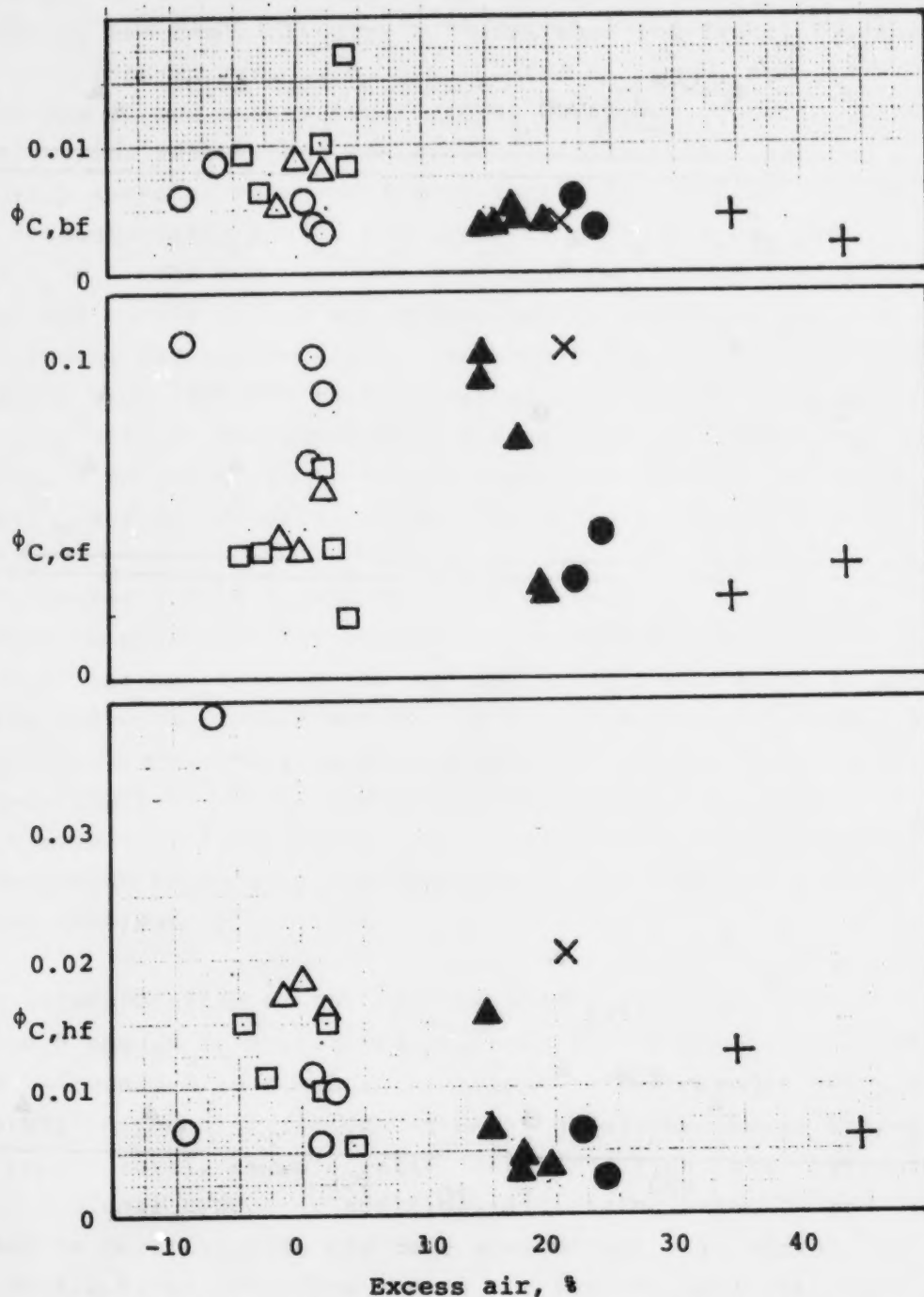


Fig. 8.6. Flux ratio of carbon loss in a product stream to carbon fed in fuel, vs. excess air level. Top, bed product; middle, cyclone product; bottom, baghouse product. Key to data: Table N.3B.

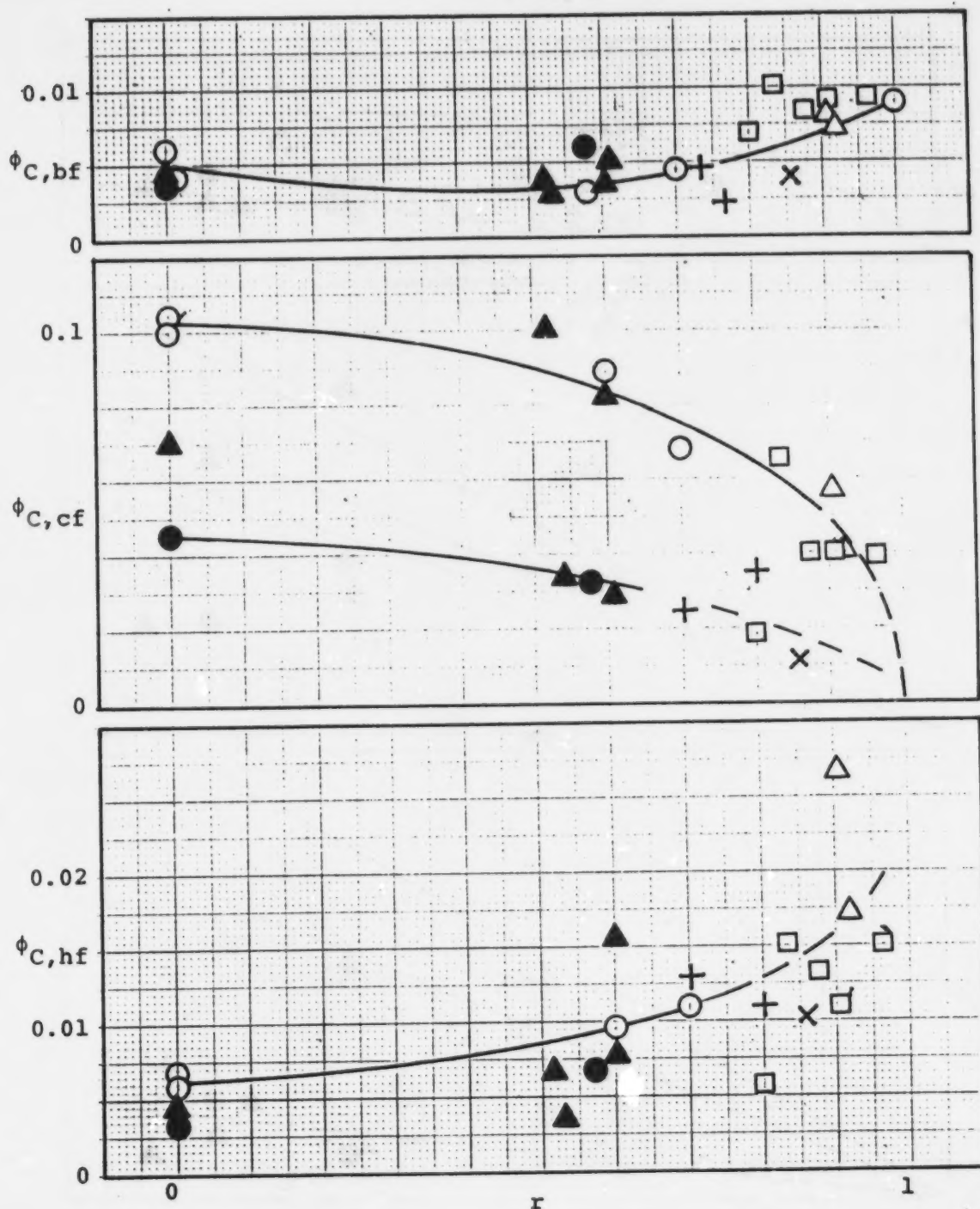


Fig. 8.7. Flux ratio of carbon loss in a product stream to carbon fed in fuel, vs. recycle fraction. Top, bed product; middle, cyclone product; bottom, baghouse product. Key to data: Table N.3B.

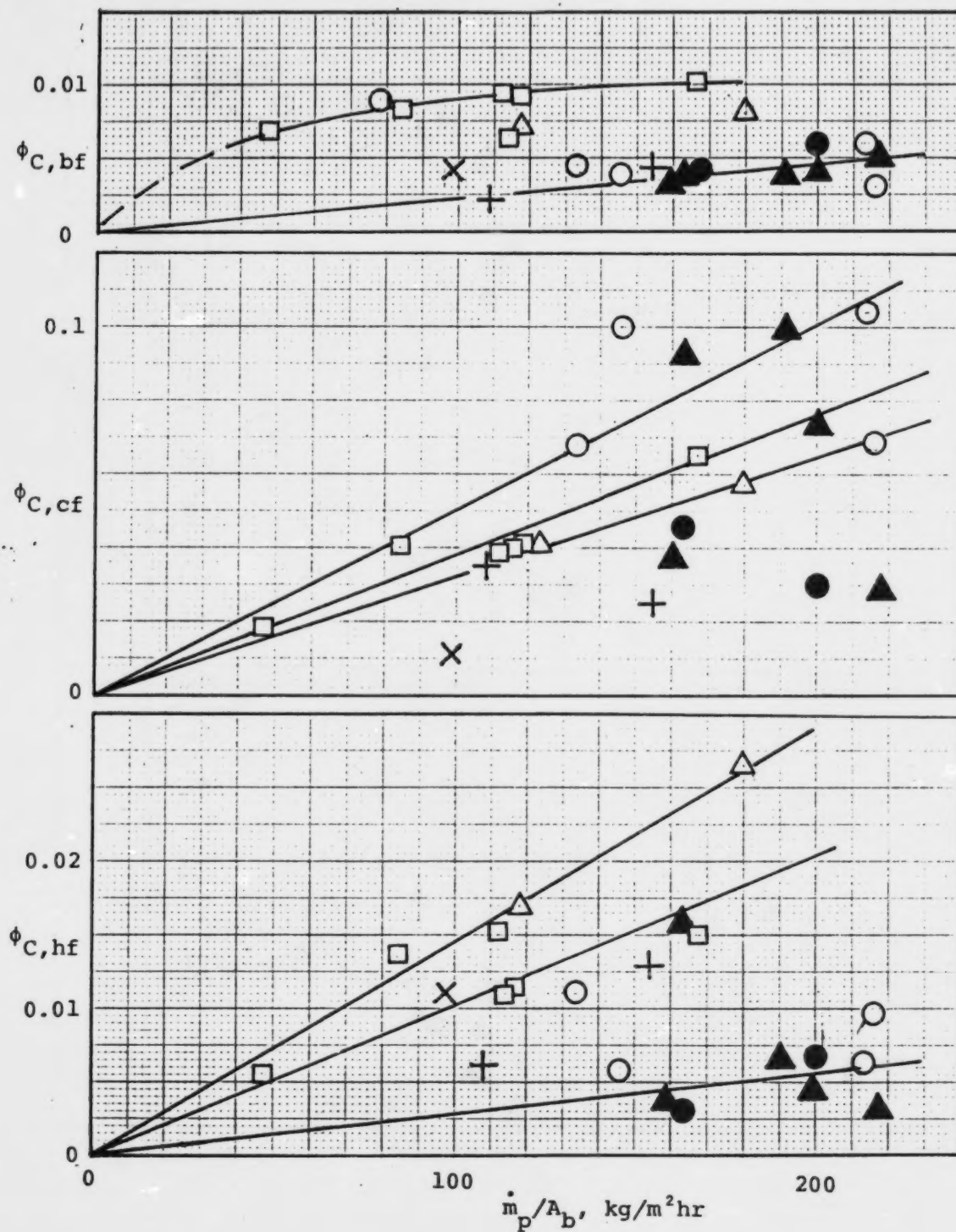


Fig. 8.8. Flux ratio of carbon loss in a product stream to carbon fed in fuel, vs. solid products flux per unit bed area. Top, bed product; middle, cyclone product; bottom, baghouse product. Key to data: Table N.3B.

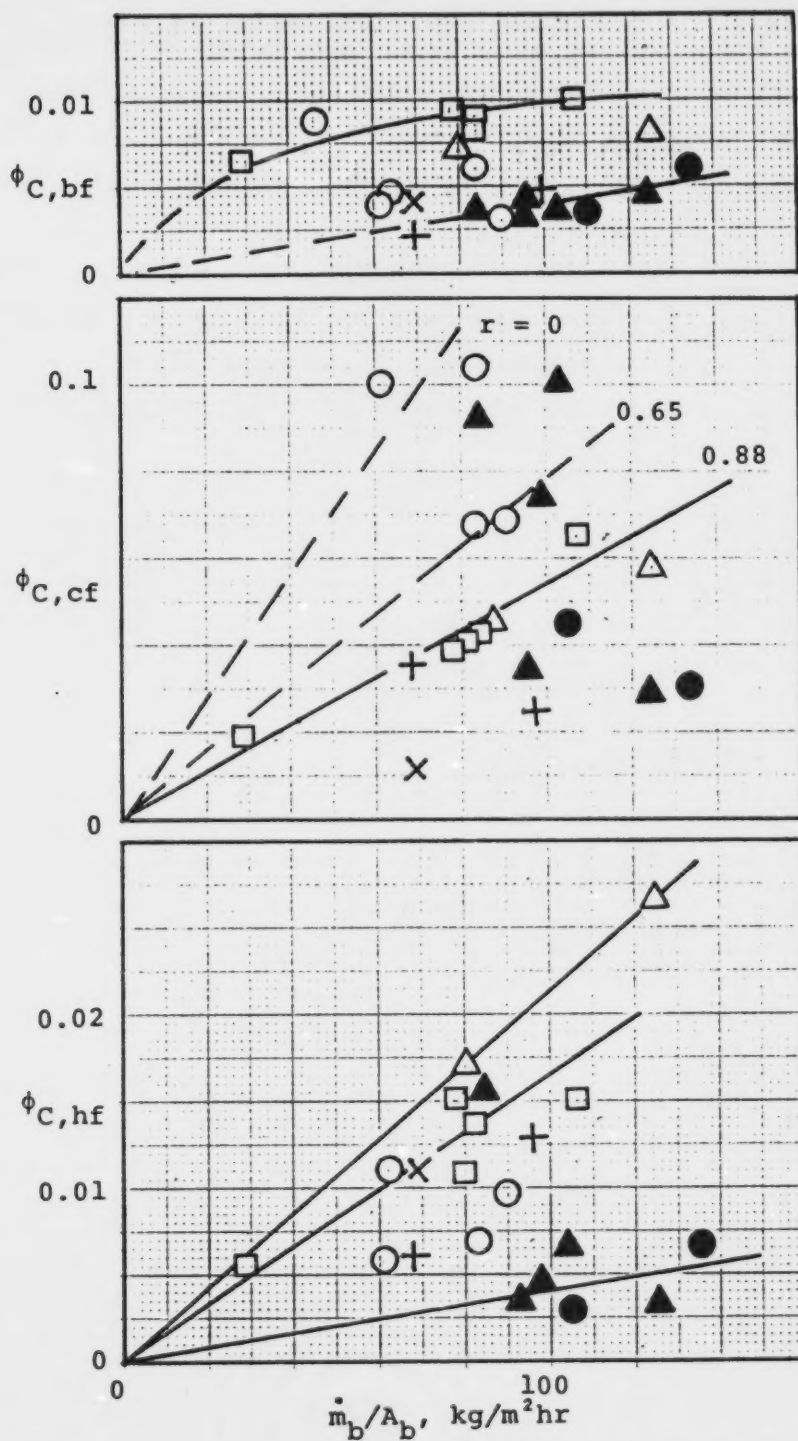


Fig. 8.9. Flux ratio of carbon loss in a product stream to carbon fed in fuel, vs. bed product flux per unit bed area. Top, bed product; middle, cyclone product; bottom, baghouse product. Key to data: Table N.3B.

mathematical models. A plausible empirical model is presented in Sect. 8.2.10.

8.2.2 Effect of bed temperature

There are indications in Fig. 8.5, from the data (shaded triangles) on the 6.6 mm Minto coal at the higher level of fluidizing velocity, that carbon loss decreases with increasing bed temperature over the range 840--950 °C. The effect is most pronounced for the cyclone and baghouse products; for these there appears to be a decrease in carbon loss on the order of 50 % per 100 °C rise in temperature.

Both here and in the following subsections of Sect. 8.2, the inferences (as indicated in Sect. 8.3.1) are not based on examination of one figure alone (even if only one is mentioned), but on Figs. 8.5--8.9 considered together.

8.2.3 Effect of excess air level

The results (open and shaded circles) on 19.3 mm Minto coal, Fig. 8.6, suggest a decrease in carbon loss with increasing excess air level. The effect is seemingly strongest for the cyclone product, in which case it is of the order of 2 % decrease per 1 % increase in excess air.

8.2.4 Effect of recycle

The results (open circles) on the 19.3 mm Minto coal at the lower level of fluidizing velocity, Fig. 8.7, indicate a significant effect of recycle on carbon loss, strongest as the recycle fraction approaches unity. The effect of increasing recycle is to increase the carbon loss in the bed and baghouse products, but to decrease it in the cyclone product. The main factors responsible for these divergent trends do not seem difficult to identify. Recycle should increase carbon burn-off and that is indeed a part of what we see in the cyclone product. Increasing recycle, however, also returns more dust to the system and so increases the dust inventory. Further, less solid product

is withdrawn from the system as cyclone product, and more of it is withdrawn as bed and baghouse product. We thus find the carbon loss with bed and baghouse product actually increasing with increasing recycle. The net effect when all carbon losses are added up, however, is that total carbon loss decreases with increasing recycle, just as expected (the total balance is discussed in Sect. 8.2.8).

8.2.5 Effect of residence time

The results (open squares) in Figs. 8.8 and 8.9 on 6.6 mm and 10.7 mm Minto coal at the lower level of fluidizing velocity seem to indicate a strong effect of residence time. As noted earlier, we consider the residence time $\tau_p = m_b/\dot{m}_p$ to be less meaningful than $\tau_b = m_b/\dot{m}_b$, because the inventory of elutriable dust in the system is usually relatively small and the residence time of such dust in the system is expected to be substantially less than τ_p or τ_b . In view of this, and because the bed product consists essentially of the same material as the bed itself, τ_b is the correct mean residence time for the bed product and the key characteristic residence time for the system as a whole. We therefore attach more significance to Fig. 8.9 than to Fig. 8.8, and indeed include Fig. 8.8 only to demonstrate that the trends indicated are quite similar.

There is some indication in the results on the cyclone product stream in Fig. 8.9 that coal particle size has rather small effect on the carbon loss. The differences between the results (all of them) for the lower level of fluidizing velocity appear to be largely due to differences in recycle. Assuming a generally linear relation between carbon loss and $\dot{m}_b/A_b$ for the cyclone product at constant recycle fraction r , and approximating the relation between the slope and $1 - r$ by a power-law function, the three curves indicated in Fig. 8.9 (middle graph) and the implied fourth (the limit at $r = 1$, of zero slope) give

$$\dot{m}_{C,c}/\dot{m}_{C,f} = 0.0013(\dot{m}_b/A_b)(1 - r)^{0.43} \quad (8.2.1)$$

Introducing $m_b = 500 A_b$, we get

$$\dot{m}_{C,c}/\dot{m}_{C,f} = 0.65 \tau_b^{-1} (1 - r)^{0.43} \quad (8.2.2)$$

This correlation applies in the narrow range of the data used to derive it: bed temperature 830--870 °C, excess air level 0--10 %, fluidizing velocity 1.6--2.2 m/s.

8.2.6 Effects of fluidizing velocity

Increasing the fluidizing velocity significantly also meant increasing the coal burning rate. At given Ca/S ratio, that in turn meant increasing the limestone feedrate. Both increases produced a shortening of solids residence time. Thus the effects of fluidizing velocity and residence time are difficult to separate. It may nevertheless be argued that fluidizing velocity should have a substantial independent effect on carbon loss, because elutriation should increase markedly with increasing fluidizing velocity. These questions are pursued further in Sect. 8.2.10.

8.2.7 The total carbon loss

Figures 8.10 and 8.11 show the total fractional fuel carbon loss $\dot{m}_{C,p}/\dot{m}_{C,f}$, $\dot{m}_{C,p} = \dot{m}_{C,b} + \dot{m}_{C,c} + \dot{m}_{C,h}$, as a function of bed temperature, excess air level, recycle fraction, and residence time. The total loss is seen to range from 2.5 % to 11.5 % of the fuel carbon. The effects of temperature, excess air, and residence time are perhaps better perceived in the component losses (considered in the preceding sections), and certainly there is nothing new to be noted here. The effect of recycle, however, is different in its totality from any of its components. We saw that recycle increases the carbon loss through bed product and baghouse product, but decreases it in the cyclone product. Now we see, as expected, that the net effect is to decrease carbon loss. The total decrease in going from zero recycle to full recycle, as indicated by the results on 19.3 mm Minto coal,

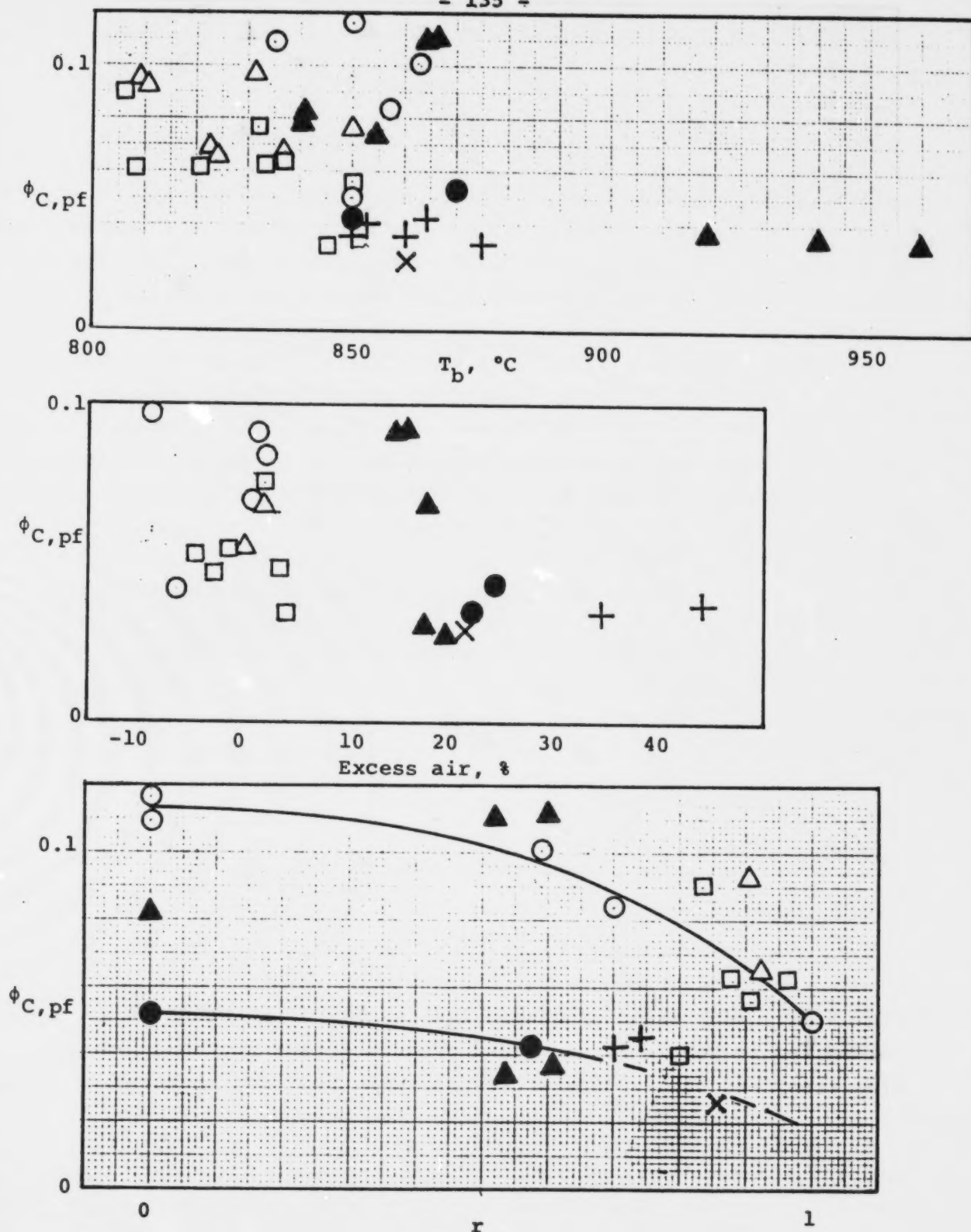


Fig. 8.10. Flux ratio of carbon loss in the solid products to carbon fed in fuel, vs. temperature (top), excess air level (centre), and recycle fraction (bottom). Key to data: Table N.3B.

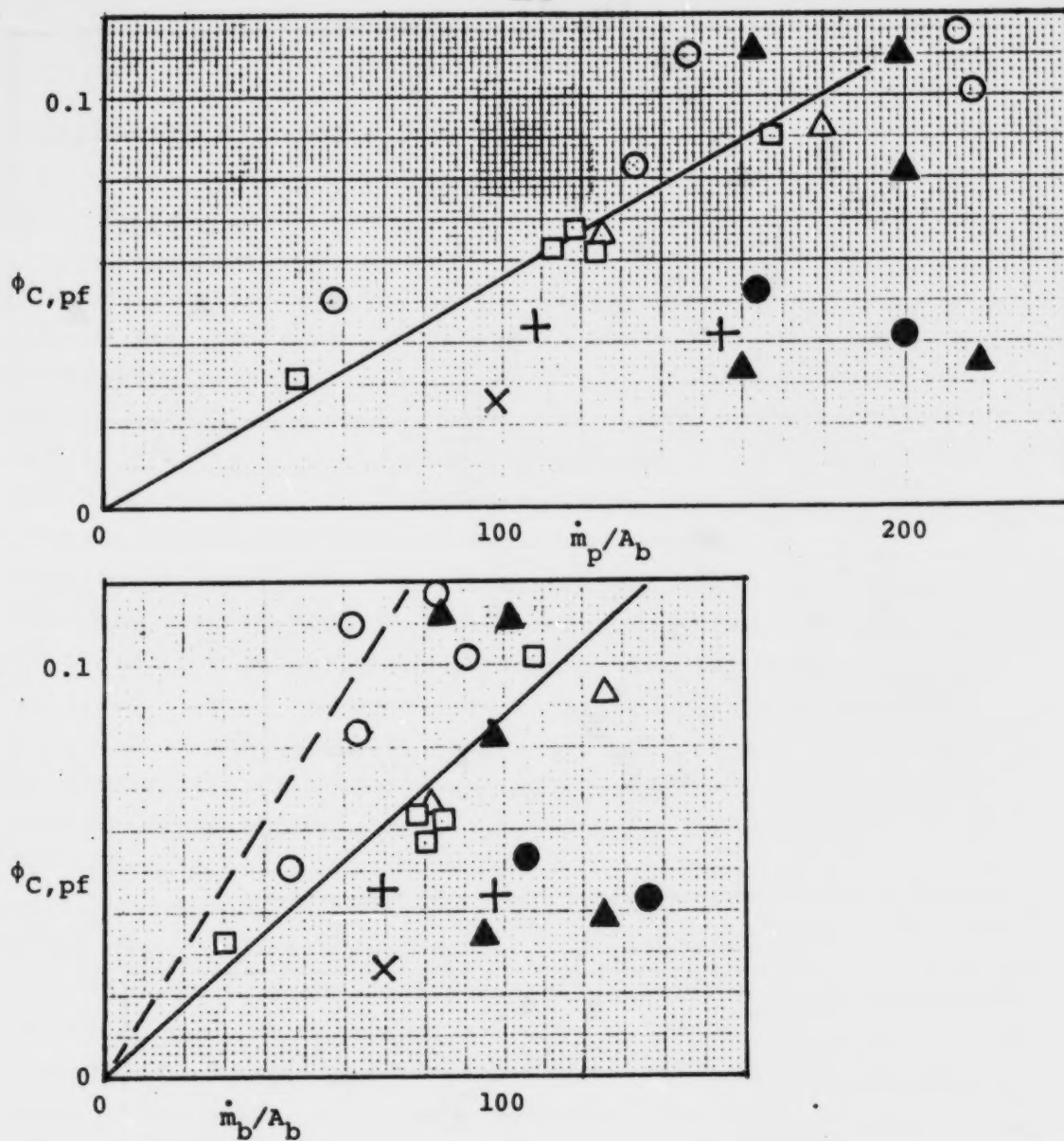


Fig. 8.11. Flux ratio of carbon loss in the solid products to carbon fed in the fuel, vs. solid product efflux per unit bed area (top) and bed product efflux per unit bed area (bottom).

may be as high as 50 %.

8.2.8 Further observations on the division of carbon loss

The division of carbon loss between the bed product, cyclone product, and baghouse product streams was briefly examined earlier (Fig. 8.4). Figures 8.12 and 8.13 show this division as a function of the single most influential variable--the recycle fraction. There is nothing new to be noted here; these figures simply serve to give a clearer quantitative impression of the division. Qualitatively, the behaviour parallels that of the solid product streams themselves, Figs. 7.5 and 7.6. The scatter, however, is much greater, and no single parameter (such as limestone particle size in the said figures) brings order to the data.

8.2.9 The cyclone/baghouse product carbon division

It was seen in Sect. 7.4 that the ratio of cyclone product to baghouse product is strongly correlated with the recycle fraction. This seems to be fairly true, too, for the ratio of the corresponding carbon fluxes, Fig. 8.14. However, the carbon flux ratios tend to be about 50 % higher.

8.2.10 An empirical model for the overhead carbon loss

The overhead carbon loss is the sum of the losses in the cyclone product and the baghouse product,

$$\dot{m}_{C,o} = \dot{m}_{C,c} + \dot{m}_{C,h} \quad (8.2.3)$$

It is that portion of the carbon leaving the combustor that is carried out by elutriation and not returned as recycle. Indications have been seen in the preceding subsections as to how this overhead loss might behave. Further indications are found in our report on elutriation (Xu, Becker and Code 1983) and in a study by Stover, Anthony, Desai and Friedrich (1981) of the

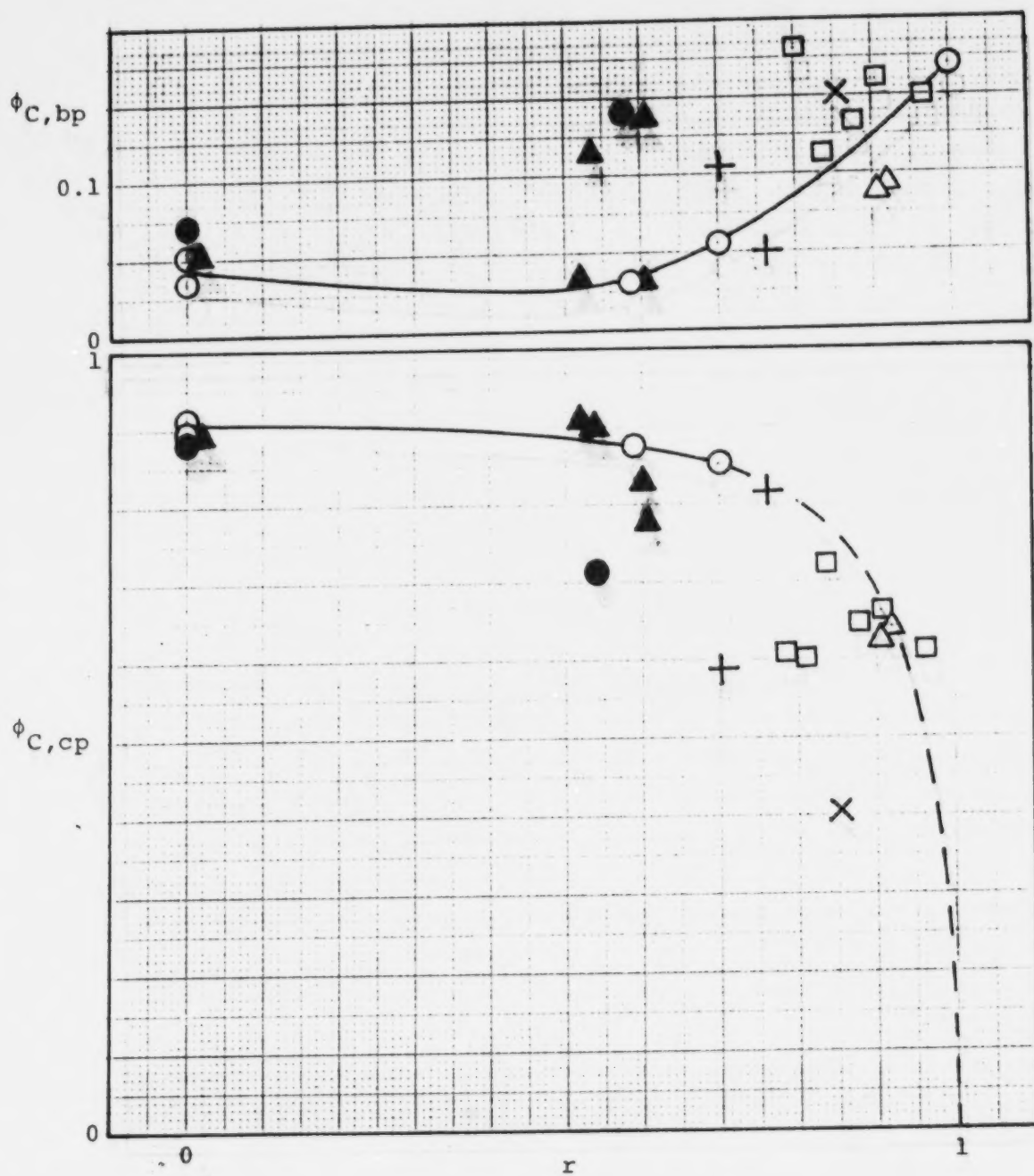


Fig. 8.12. Flux ratio of carbon loss in bed product (top) and cyclone product (bottom) to total loss in solid products, vs. recycle fraction. Key to data: Table N.3B.

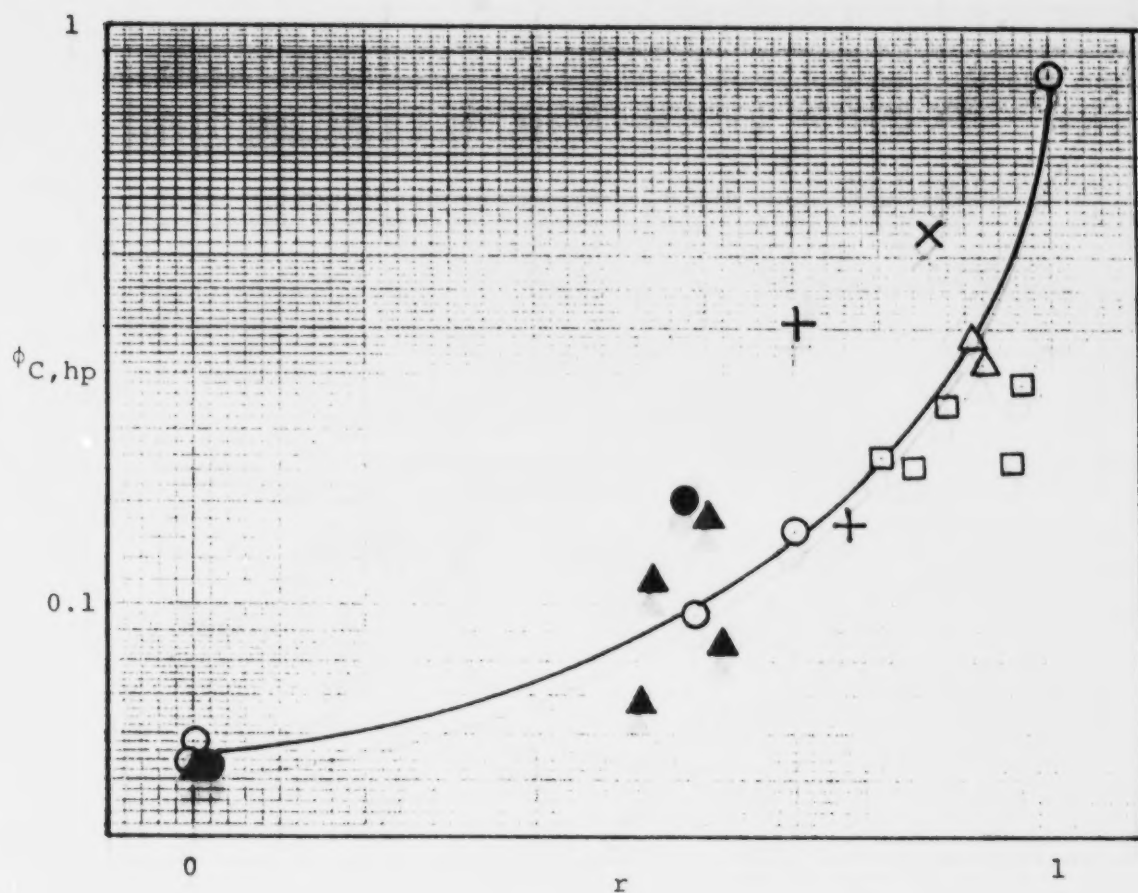


Fig. 8.13. Flux ratio of carbon loss in the baghouse product to total loss in solid products, vs. recycle fraction. Key to data: Table N.3B.

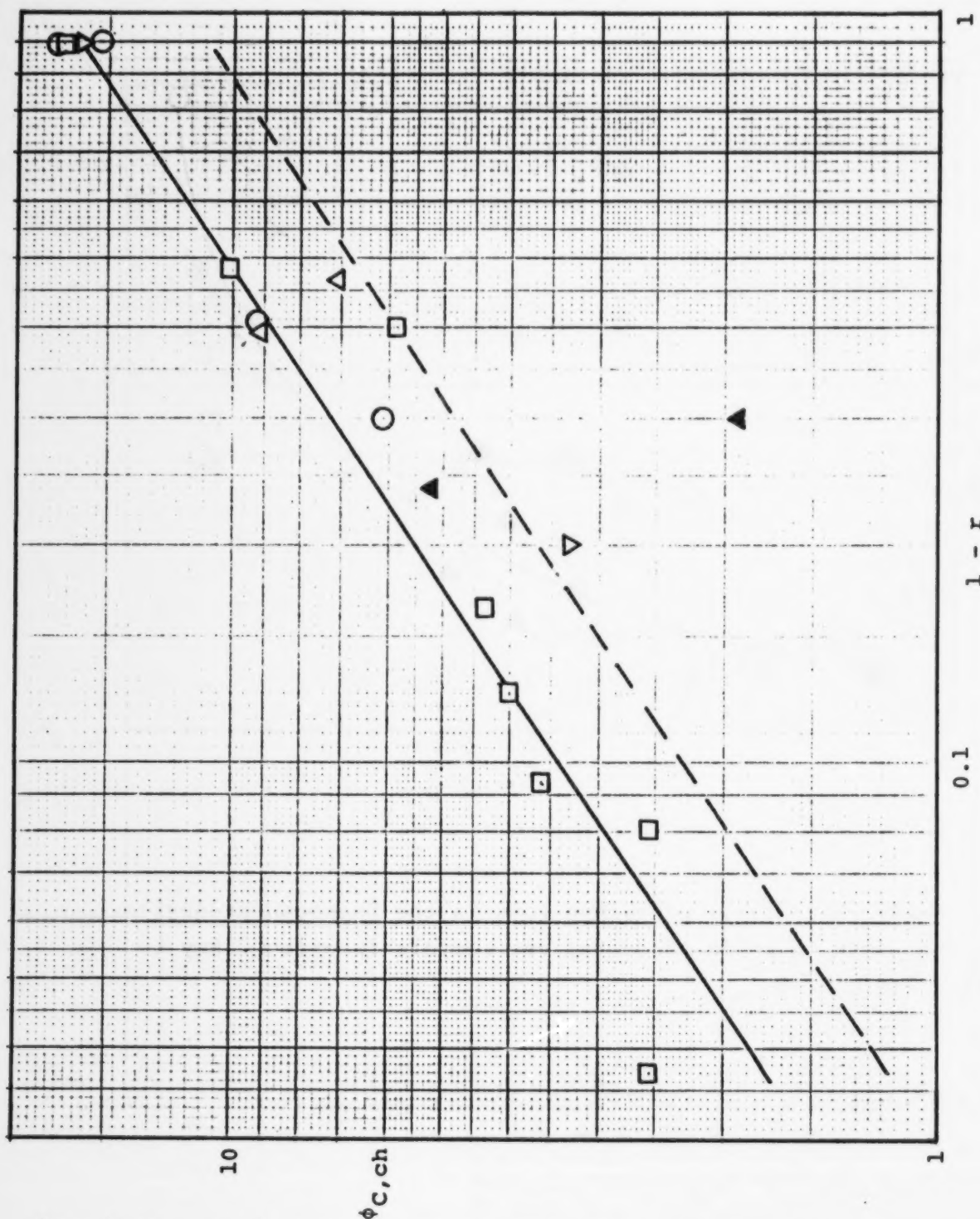


Fig. 8.14. Flux ratio of carbon loss in cyclone product to loss in baghouse product, vs. the non-recycle fraction $1 - r$. Key to data: Table N.3A. Compare with ϕ_{ch} in Fig. 7.7; to a close approximation, $\phi_{C, ch} = 1.5 \phi_{ch}$. The dashed line represents (7.4.1); the full line represents ordinate values 1.5 times larger.

burning of Minto coal in a smaller fluidized-bed combustor without limestone addition and without recycle (the findings of this significant report are fully discussed in Sect. 18). After examining all the factors that might be influential (bed temperature, excess air, recycle, solids residence time, fluidizing velocity, limestone particle diameter, and coal particle diameter), and possible forms by which those influences might be suitably expressed over the normal operating range, we narrowed consideration to the following model:

$$y = a_0 + a_1 x_1 + a_2 x_2 + a_3 x_3 + a_4 x_4 + a_5 x_5, \quad (8.2.4)$$

$$y = \ln(\dot{m}_{C,o}/\dot{m}_{C,f})$$

$$x_1 = \ln v$$

$$x_2 = \ln \phi_a$$

$$x_3 = \ln \hat{D}_p$$

$$x_4 = 1/T_b$$

$$x_5 = r,$$

where $\phi_a = \dot{m}_a/\dot{m}_{a,st} = 1 + \epsilon_a$ is the "mixture strength" of Stover, et al., $\dot{m}_{a,st}$ is the stoichiometric (theoretical) air requirement for complete combustion, and T_b is the absolute bed temperature. The model may also be written in the equivalent power-law form

$$\dot{m}_{C,o} = C_o \dot{m}_{C,f} v^{a_1} \phi_a^{a_2} \hat{D}_p^{a_3} \exp(a_4/T_b) \exp(a_5 r), \quad (8.2.5)$$

where $C_o = \ln a_o$. A model like this in v , ϕ_a and T_b was used by Stover, et al. Models of this type, with power law

dependencies on gas velocity and characteristic diameter and an Arrhenius-like temperature dependency, have been very successfully used in our laboratories to describe the burning of wood (Phillips and Becker 1982). The exponential dependency on recycle fraction appears likely to give a reasonably accurate representation of that effect while not complicating the statistical fitting problem (we have actually no basis for preferring any more complicated or mathematically awkward form).

The equation (8.2.4) was statistically fitted to our data by multiple linear regression analysis. One data set $\{y, x_1, x_2, \dots\}$ representing steady state conditions was entered for each run, Table 8.1. The computed values of the coefficients a_n are presented in Table 8.2. Also given are some of the statistics on goodness of fit: the standard errors s_a of the a_n , the t statistics for the a_n , the standard error s_y of the estimate of y , and the multiple correlation coefficient R . Results are shown first for the whole data set and then for three subsets.

- o the subset excluding runs 830126, 830202 and 830215. The overhead carbon loss in these runs was significantly high but there was nothing unusual that might explain this in the values of the considered independent variables (the x_n) or those excluded by preliminary screening as uninfluential (coal particle diameter and solids residence time):
- o the subset excluding the Devco runs. This introduces coal type as a parameter.
- o the subset excluding both the said Minto runs with exceptional carbon loss and the Devco runs.

We have the following observations on these results:

1. The coefficient of $\ln \hat{D}_p$ is determined best and varies little, 0.72--0.82, between the four sets.

Table 8.1. Data on the fractional overhead carbon loss

$$\phi_{C,of} \equiv \dot{m}_{C,o} / \dot{m}_{C,f}$$

Run	v m/s	ϕ_a	$\hat{D}_p, \text{mm}$		T_b °C	r	$\phi_{C,of}$
			Stone	Coal			
830106	1.69	1.04	1.98	10.7	845	.80	.0241
830112	1.68	.95	.78	10.7	821	.96	.0532
830124	1.86	.97	.78	10.7	849	.91	.0496
830126	1.78	1.02	.78	10.7	832	.84	.0677
830127	1.79	1.03	.78	10.7	833	.88	.0534
830201	1.90	1.00	.78	6.6	823	.92	.0596
830202	1.90	1.02	.78	6.6	809	.91	.0838
830205	2.77	1.15	.78	6.6	866	.60	.1074
830216	2.76	1.12	.78	6.6	855	.52	.0706
830217	2.68	1.18	.78	6.6	841	0	.0772
830218	2.77	1.24	1.98	19.3	870	0	.0482
830224	2.68	1.22	1.72	19.3	849	.57	.0363
830225	2.98	1.19	1.72	6.6	959	.54	.0291
830226	2.93	1.17	1.72	6.6	941	.61	.0304
830322	1.82	1.01	.49	19.3	835	0	.1057
830324	2.06	1.01	.49	19.3	857	.70	.0788
830329	2.25	1.02	.49	19.3	862	.59	.0980
830330	2.02	.91	.49	19.3	850	0	.1107
830309	1.90	1.21	1.72	6.6	860	.86	.0212
830315	2.73	1.34	1.72	6.6	864	.70	.0369
830316	2.67	1.44	1.72	6.6	852	.76	.0374

Table 8.2. Correlation of overhead carbon loss: results of regression analysis on the form $y = \sum a_n x_n$, $y = \ln(\dot{m}_{C,o}/\dot{m}_{C,f})$.

x_n	1	$\ln v$	$\ln \phi_a$	$\ln \hat{D}_p$	$1/T_b$	r
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All runs: $s_y = 0.16$, $R = 0.923$

a_n	-9.86	1.23	-0.69	-0.78	6900	-0.23
s_a	2.13	0.40	0.71	0.13	2200	0.13
t	-4.6	3.1	-1.0	-6.1	3.2	-1.8

Three Minto runs* excluded: $s_y = 0.10$, $R = 0.969$

a_n	-8.00	0.88	-0.46	-0.74	5200	-0.36
s_a	1.50	0.28	0.48	0.08	1500	0.09
t	-5.4	3.2	-1.0	-8.8	3.4	-4.1

Devco and three Minto* runs excluded: $s_y = 0.09$, $R = 0.978$

a_n	-6.89	0.66	-0.52	-0.72	4100	-0.40
s_a	1.42	0.33	0.74	0.09	1400	0.08
t	-4.8	2.0	-0.7	-8.4	2.9	-5.1

Devco runs excluded: $s_y = 0.17$, $R = 0.907$

a_n	-8.61	0.85	0.04	-0.82	5800	-0.23
s_a	2.53	0.59	1.36	0.16	2500	0.14
t	-3.4	1.4	0.0	-5.3	2.3	-1.7

*Runs 830126, 830202 and 830215

2. The coefficient of $\ln \phi_a$ is the least well determined, and the significance of this factor is indeed doubtful.

3. The coefficients of $\ln v$ and $1/T_b$ are substantially affected by omission of the exceptional Minto runs.

4. Examination of the results indicates no significant difference between the Devco and Minto coals.

5. In the study of elutriation (Xu, Becker and Code 1983), the importance of a parameter $D_p/v^{1.163}$ was very precisely established, where D_p is the diameter of bed particles of a given size. The mean limestone particle size $\hat{D}_p$ here generally closely approximates the mean size of bed particles (it is around 10 % larger). There is a consistency then, and it is well within the uncertainty of the results in Table 8.2, in supposing that here a parameter $\hat{D}_p/v^{1.63}$ is involved, where $\hat{D}_p$ is the mean limestone particle diameter.

Overall, it appears to us that the correlation best representing the dominant trend of the results is the second one in Table 8.2, namely that based on exclusion of the three exceptional Minto runs. We therefore take the coefficients of $\ln \hat{D}_p$, $1/T_b$ and r to be respectively - 0.74, 5200 and - 0.36. The coefficient of $\ln \phi_a$, which is very uncertain, is rounded to 0.5. Finally, following the fifth of the above observations, we take the coefficient of $\ln v$ to be 1.163 times that of $\ln \hat{D}_p$, thus 0.86. The constant a_0 is virtually unchanged by these adjustments. The resulting correlation (written in the power-law form) is

$$\dot{m}_{C,o} = 0.000335 \dot{m}_{C,f} v^{0.86} \hat{D}_p^{-0.74} \phi_a^{-0.5} \exp(-0.36 r) \exp(5200/T_b), \quad (8.2.6)$$

where v is in m/s and $\hat{D}_p$ in millimetres. The standard error in the prediction of the constant 0.000335 (with the three exceptional Minto runs excluded) is 0.000029, or 8.7 %. The run-by-run values of

$$C_o = (\dot{m}_{C,o}/\dot{m}_{C,f}) v^{-0.86} \hat{D}_p^{0.74} \phi_a^{0.5} \exp(0.36 r) \exp(-5200/T_b) \quad (8.2.7)$$

are shown as a function of the recycle fraction r in Fig. 8.15 in which the data symbols distinguish between the coal particle diameters, the coal types, and the levels of fluidizing velocity. The figure illustrates the good quality of the correlation for the "normal" runs and the magnitude of the abnormality for the three exceptional ones. We have at present no explanation for the exceptional cases except for the possibility that they represent rare steady-state conditions in a situation where several steady states may occur (see Sect. 13).

The correlation indicates that the maximum effect of increasing recycle, going from $r = 0$ to $r = 1$, is to decrease the overhead carbon loss by 30 %. This is significant but quite unspectacular. If this is the total benefit of recycle, it may not be worth the cost of installing and operating a recycle system. It should be noted that the data set, Table 8.1, excludes the few runs (all made at the beginning of our program) with total recycle, $r = 1$. Because of problems at the time, baghouse samples for chemical analysis were lost and so the overhead carbon loss could not be determined. We are thus here relying on extrapolation to $r = 1$ from very near that limit.

Highly significant is the fact that coal particle diameter has no detectable effect. The implications of this, beyond the obvious and practically important one that carbon loss is virtually unaffected by coal particle size, are examined in Sect. 18.

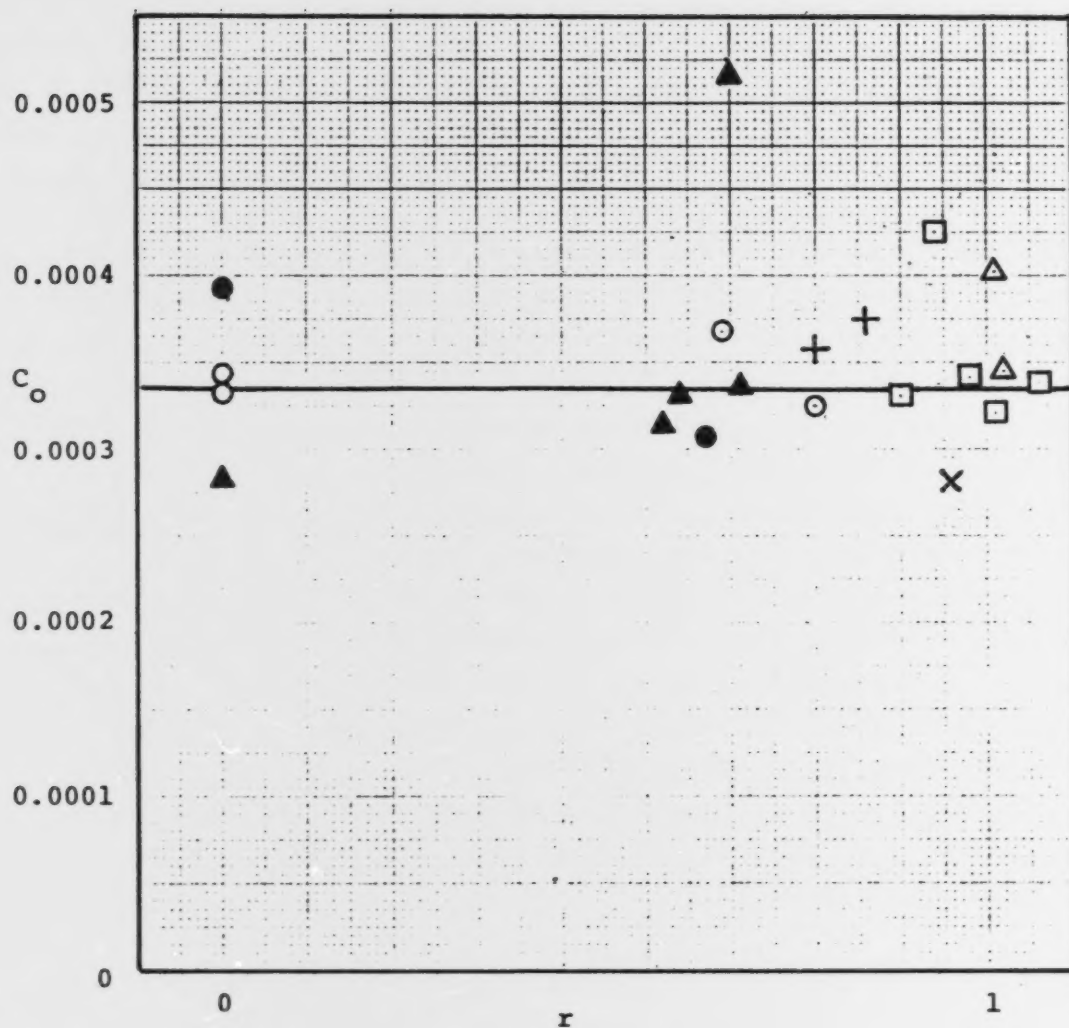


Fig. 8.15. The coefficient C_o in the correlation of overhead carbon loss (8.2.6), calculated for the individual data points, vs. recycle fraction. The statistically fitted value is $C_o = 0.000335$. Key to data: Table N.3B.

Very significant, too, is the finding that overhead carbon loss decreases with increasing limestone particle diameter. Implications of this are also discussed in Sect. 18.

The fact that solids residence time is not needed in the correlation has several possible explanations. First, fluidizing velocity and solids feedrate are considerably correlated in the mode of operation of present interest (low to moderate excess air, as in normal power and heating plants). Thus the fluidizing velocity serves to largely cover any effects of residence time. Most likely to be important, however, is the fact that the residence time that is really at issue here is that of significantly carbon-containing elutriated particles. This is likely very different from the mean residence time of bed solids (τ_b or τ_p), and probably behaves altogether differently. As far as we can tell, the correlation as now constituted in (8.2.6) contains all that is necessary in this regard, although a model physically explaining it remains to be developed. If this is so, then the remarks in Sect. 8.2.5 on effects of residence time are of decidedly limited validity.

The overhead carbon loss typically accounts for 80--97 % of the total carbon loss from the combustor. Thus ability to predict it practically amounts to an ability to predict the total loss.

8.3 Combustion efficiency and carbon burn-up

Combustion efficiency is best defined as a thermal measure of completeness of burning, e.g., the actual chemical heat release divided by the "theoretical" heat release occurring if the fuel were completely oxidized. In the present case, the thermal measure differs little from the carbon burnup efficiency,

$$\eta_C = 100(1 - \dot{m}_{C,p}/\dot{m}_{C,f}) \quad (8.3.1)$$

Figures 8.10 and 8.11 show that this carbon burnup efficiency ranged from 88.5 % to 97.5 %. Since the overhead carbon loss $\dot{m}_{C,o}$ accounts for nearly all of the total carbon loss $\dot{m}_{C,p}$, the correlation presented in Sect. 8.2.10 practically suffices for the prediction of $\dot{m}_{C,p}$.

8.4 The input/output carbon balance

Carbon enters the system as fuel carbon in the coal, carbonate carbon in the limestone, and CO_2 in the air. The last is quite negligible here. Carbon leaves as CO_2 and CO in the exhaust gases, as a component of char in the solid products, and as carbonate carbon due to uncalcined limestone in the solid products. The last of these should be small or negligible; in any case, our carbon analysis of the solid products did not distinguish between carbonate carbon and char carbon and simply measured their sum.

Figure 8.16 shows the results on the total carbon balance--carbon efflux vs. carbon influx. The balance is satisfactory, the deviations being consistent with the net standard deviation to be expected as a result of the uncertainties in the component carbon fluxes.

8.5 Solids carbon analyses at intermediate points

Solids samples were taken from the bed with a water-cooled sampling probe (described in QFBC.TR.82.1) at ports 244 mm and 569 mm above the distributor plate, on the side opposite the recycle injection port, horizontally near the middle of the side. The chemical analyses show, Fig. 8.17, that the carbon content was typically 50 % higher than in the bed product which was taken by gravity overflow from the top of the bed 1190 mm above the distributor plate. We are not sure why this was. It appears most likely that the probe was catching carbon-enriched solids emanating from the recycle injection port. There is no correlation of the effect with coal particle size; so non-uniformities emanating from the coal feed port do not seem to be

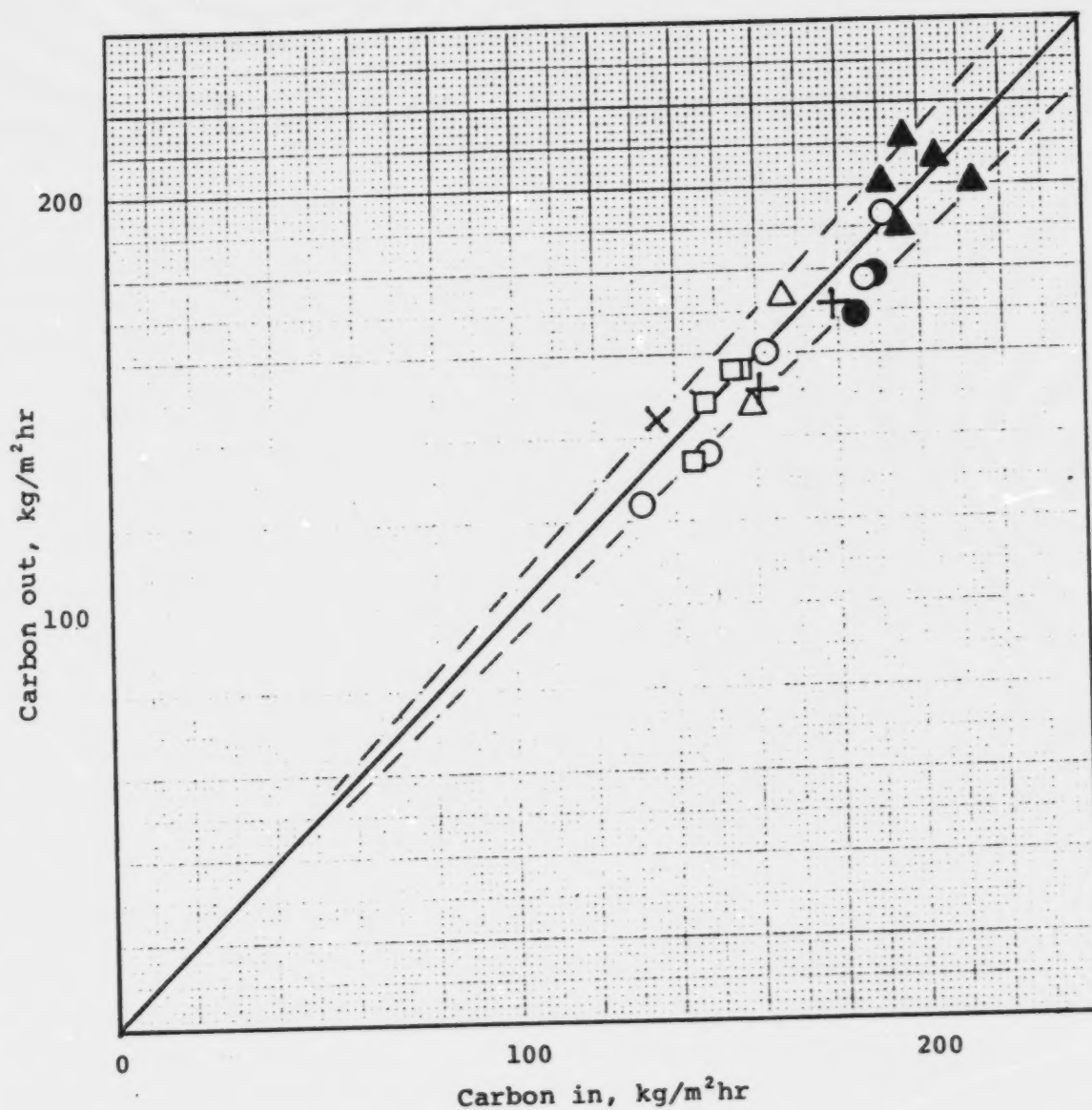


Fig. 8.16. The total carbon balance: carbon flux per unit bed area in product streams vs. that in feed streams. The dashed lines represent $\pm 7\%$ deviation. Key to data: Table N.3B.

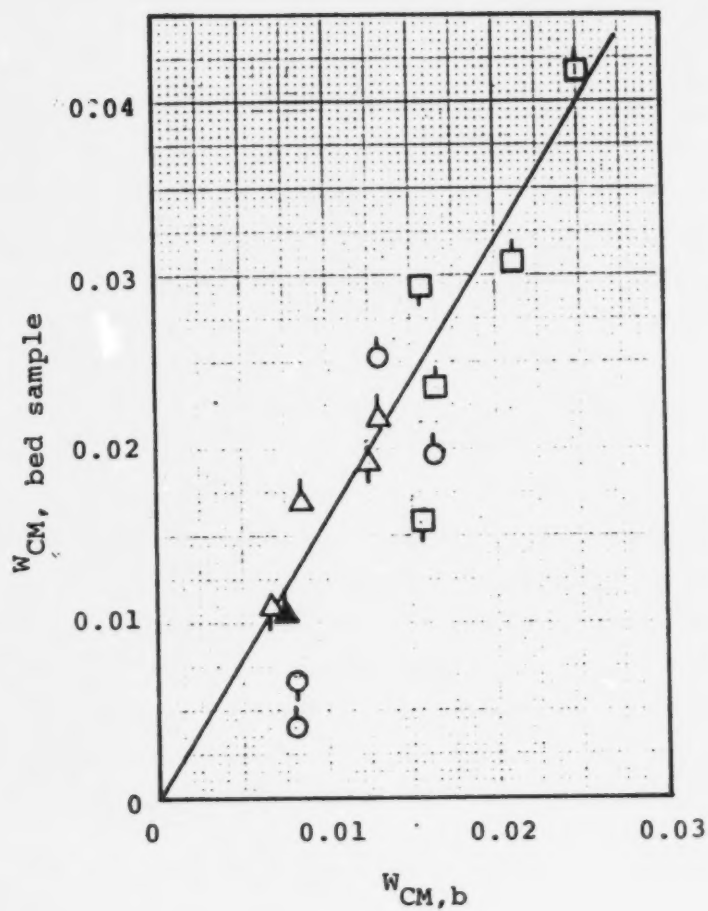


Fig. 8.17. Combustible matter in solids sample from bed vs. that in bed product. Sampling ports, Fig. 14.2: ○□△, no. 3; ○□△▲, no. 34. Coals: ○□△, Minto; ▲, Devco. Coal particle dia. D_p , mm: ○, 19.3; □, 10.7; △▲, 6.6. Slope of curve through data for 6.6 and 10.7 mm coals: 1.5.

implicated.

Elutriating particles were sampled with a dust sampling probe (described in QFBC.TR.82.1) from the gases at the combustor exit. Figure 8.18 shows that the carbon contents agreed fairly well with the values calculated from the analyses of the cyclone and baghouse products. The latter are given by

$$w_{C,x} = \frac{1}{\dot{m}_e} \left(\frac{\dot{m}_c w_{C,c}}{1 - r} + \dot{m}_h w_{C,h} \right) \quad (8.5.1)$$

The level of scatter of the data seen in these comparisons is probably not unreasonable. The local samples were taken with the probes over relatively short periods of time, typically 2 minutes for the in-bed probe and 30 minutes for the exhaust gas probe. Bed, cyclone and baghouse products, on the other hand, were collected for 90 minutes or more and then sampled, so fluctuations in composition were more strongly damped.

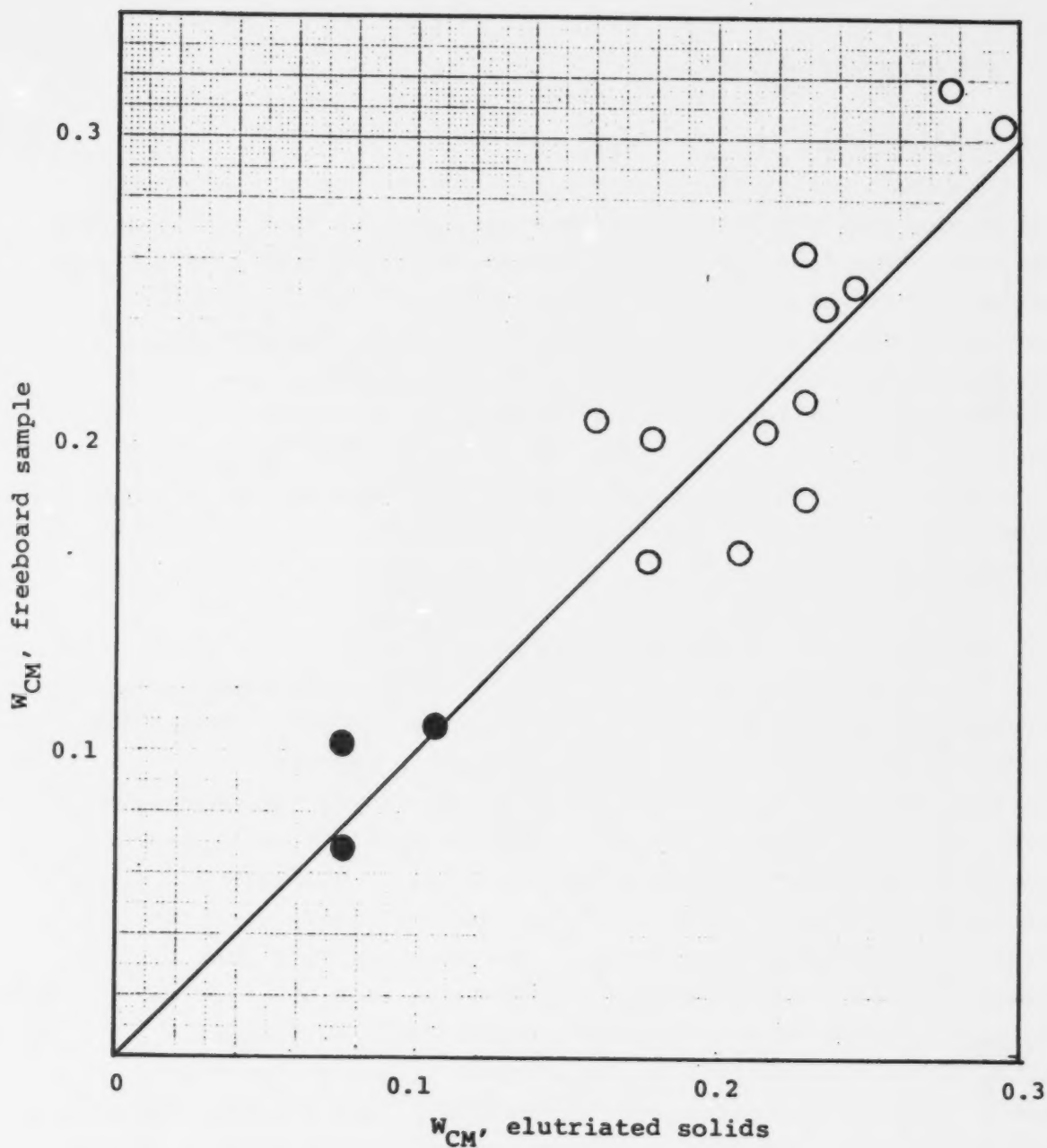


Fig. 8.18. Combustible matter in dust sample from top of freeboard vs. that in the elutriated solids (baghouse product, cyclone product, and recycle). Coals: ○, Minto; ●, Devco.

9. THE HYDROGEN BALANCE

9.1 Introduction

Hydrogen enters the system with water in the air (humidity) and in the coal and limestone (moisture), and as fuel hydrogen in the coal. The fuel hydrogen is largely burned off in the bed, ending as product water in the exhaust gases. The unburned portion is chemically bound with carbon in the unburned char. The fuel H/C atomic ratio in the char is expected to be significantly smaller than that in the coal.

The fuel hydrogen content of both the Minto and Devco coals, Sect. 4.1, was about 4.1 % d.b. The atomic ratio (fuel H)/(fuel C) was about 0.70.

9.2 The H/C ratio in the solid products

Figure 9.1 shows the apparent fuel hydrogen content as a function of the apparent fuel carbon content for the three solid product streams (bed, cyclone and baghouse). We call the hydrogen contents "apparent" fuel hydrogen because we cannot be quite sure that the hydrogen measured was wholly fuel hydrogen. Some of it may have been water hydrogen due to moisture unavoidably picked up by exposure to the atmosphere at certain steps in the process of analysis. The analyses were done by a commercial analyst and we have become aware that the type of analyzer he used is not immune to such error. We further characterize the measured values of fuel carbon as "apparent" because some of the carbon may in fact have been carbonate carbon from uncalcined limestone. However, we believe this error in the fuel carbon to be very small, and likely negligible.

The range of atomic H/C ratios in Fig. 9.1 for the cyclone and baghouse products, 0.1--0.5, compared to 0.70 in the coal, is plausible. A great majority of the data for the cyclone product are in the range 0.1--0.25, and most of those on the baghouse

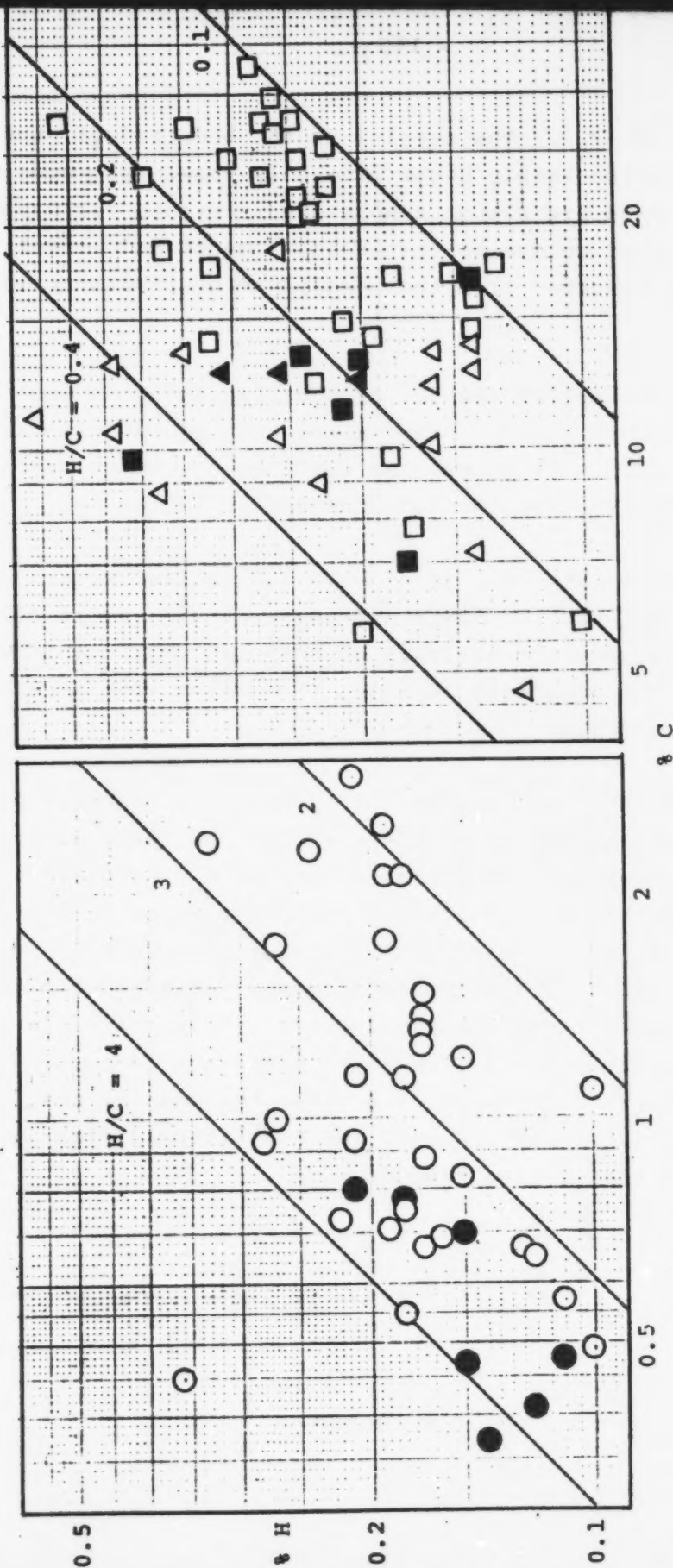


Fig. 9.1. Hydrogen content of solid products vs. carbon content.
 Product: $\bigcirc$, bed; $\square$, cyclone; $\triangle$, baghouse. Coals:
 $\bigcirc$, Minto; $\bullet$, Devco. The 45° lines represent
 fixed molar H/C ratios. In the Minto coal, $H/C = 0.060 \text{ kg/kg} = 0.72 \text{ mol/mol}$. In the Devco, $H/C = 0.063 \text{ kg/kg} = 0.76 \text{ mol/mol}$.

product are 0.15--0.3. The results on the bed product, however, $H/C = 1--4$, are impossibly high. The only reasonable conclusion is that these data are almost completely in error, and worthless, and that this must be due to the problem (mentioned above) with moisture being absorbed from the atmosphere by samples being analyzed.

The char content of the bed product, as shown in Sect. 8, was relatively very low, and thus absorption of the same amount of water by a sample of given size would be very much more serious than for the cyclone and baghouse products.

To avoid problems of this kind in the future, we are attempting to get our own C/H/N analyzer, of the latest model best suited for accurate analysis of coal and char. This will also minimize the amount of handling of samples with possible exposure to humid air.

Taking the H/C analysis of the cyclone and baghouse products to be reasonably reliable, we may then ask how these are affected by the system operating conditions. Figure 9.2 suggests a dominant effect of bed temperature for the cyclone product, with $H/C = 0.10--0.15$ up to nearly 850 °C and rising to 0.25--0.5 at 900--950 °C. Figure 9.3 shows more complex behaviour in the baghouse product with an apparent effect of coal particle size in that the H/C ratio with 19.3 mm Minto coal (the largest particles) is very high, up to 0.6. For the smaller coal particle sizes (10.7 and 6.6 mm), the H/C ratios differ little from those for the cyclone product.

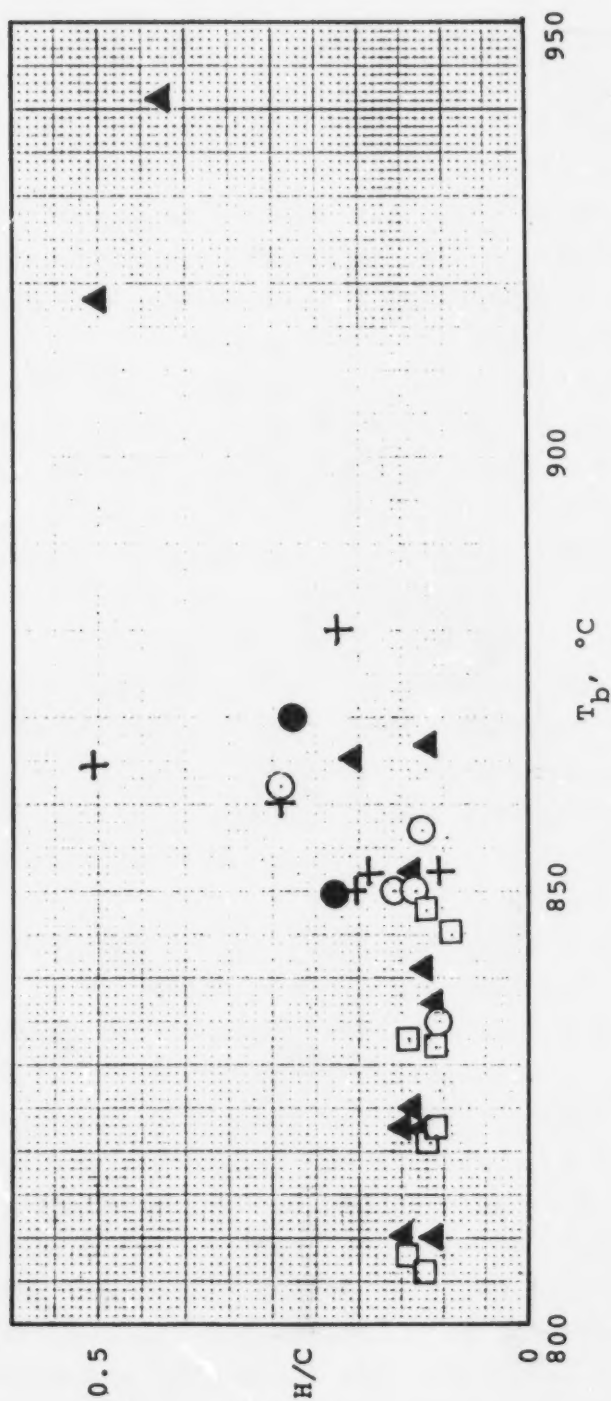


Fig. 9.2. The molar H/C ratio of the cyclone product, vs. bed temperature. Key to data: Table N.3B.

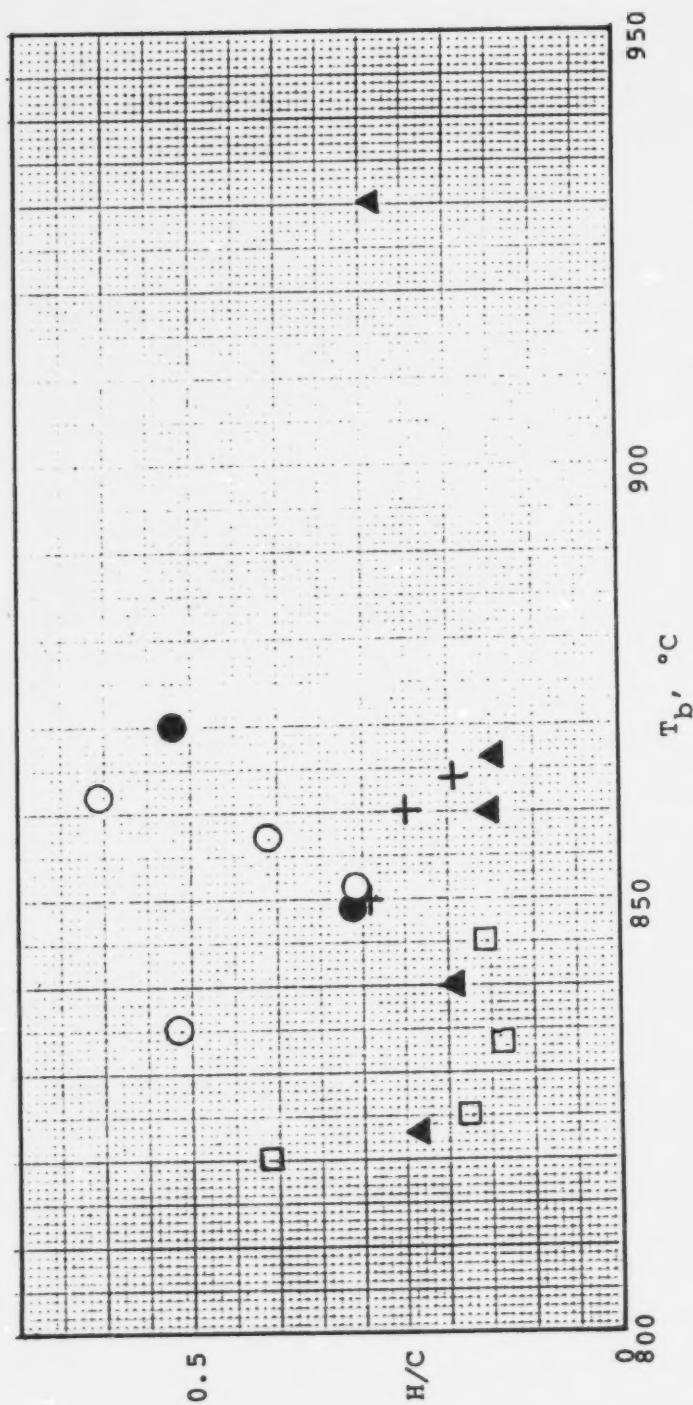


Fig. 9.3. The molar H/C ratio of the baghouse product, vs. bed temperature. See Fig. 9.1 for key to data.

10. THE FUEL NITROGEN BALANCE

10.1 Introduction

Nitrogen enters the system as fuel nitrogen and air nitrogen. It leaves as gaseous nitrogen, as gaseous nitrogen oxides, and as fixed nitrogen bound up in the char. At the low operating temperatures occurring in fluidized bed combustion (as compared to pulverized-fuel flames), the oxidation of air nitrogen must be miniscule. Any production of nitrogen oxides is therefore ascribable to reactions of fuel nitrogen. Hence attention here is focussed on the fate of the fuel nitrogen.

10.2 Nitrogen oxides emission levels

Figure 10.1 shows the NO_x level in the exhaust gas (dry basis) as a function of bed temperature. The choice of bed temperature as the "independent variable" has no special significance; it is simply convenient. The range of the NO_x is 170--415 ppm. Our instrument gave two nitrogen oxides analyses: NO and NO_x . The NO levels were only slightly lower than the NO_x , generally by 0--35 ppm, usually by about 20 ppm.

10.3 The N/C ratio in the solid products

Figure 10.2 shows the nitrogen content of the solid products (bed, cyclone and baghouse) as a function of the carbon content. A fairly high linear correlation exists in the cyclone and baghouse products, Table 10.1, and it is noteworthy that the values are nearly the same as in the raw coal. However, the results on the bed product, as with the hydrogen (Sect. 9.2) show great scatter, and unbelievably high N/C ratios relative to the raw coal. We are here led to wonder if our commercial analyst was not experiencing some difficulty simply associated with the very small amount of char in the bed product which might be a problem when using (as he was) very small samples of fixed size.

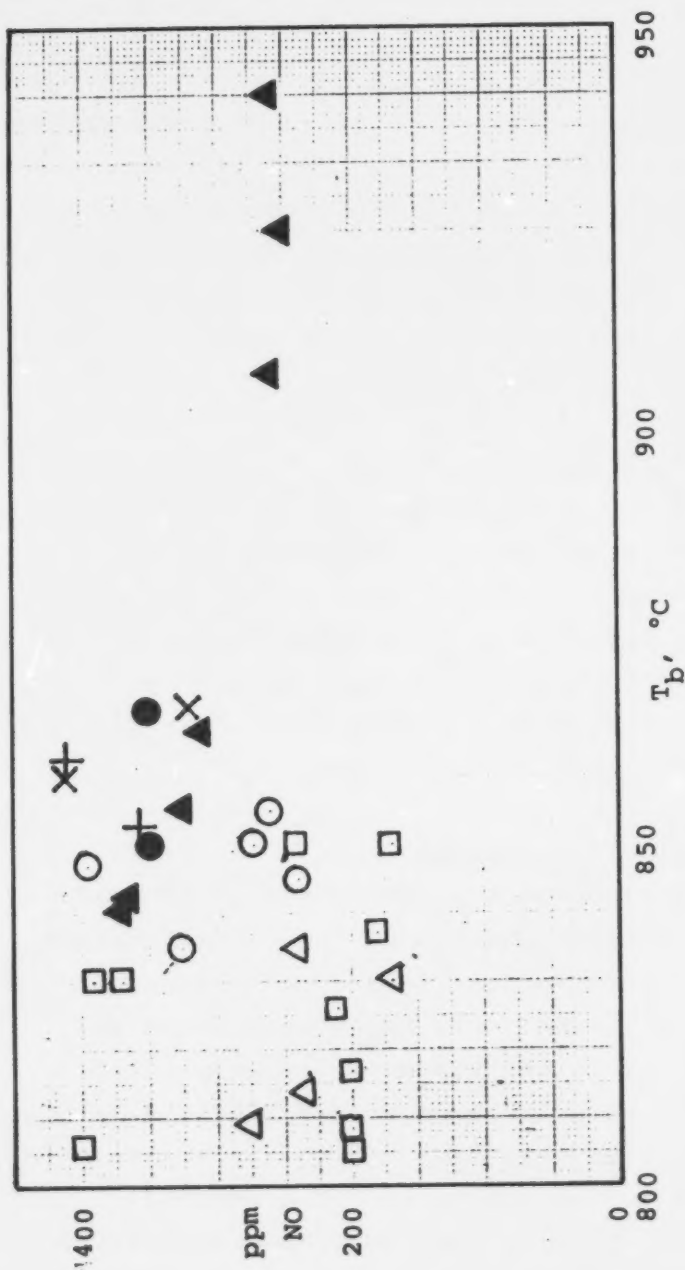


Fig. 10.1. The NO_x level (ppm by volume, dry basis) in the gas at combustor exit, vs. bed temperature. Key to data: Table N.3B.

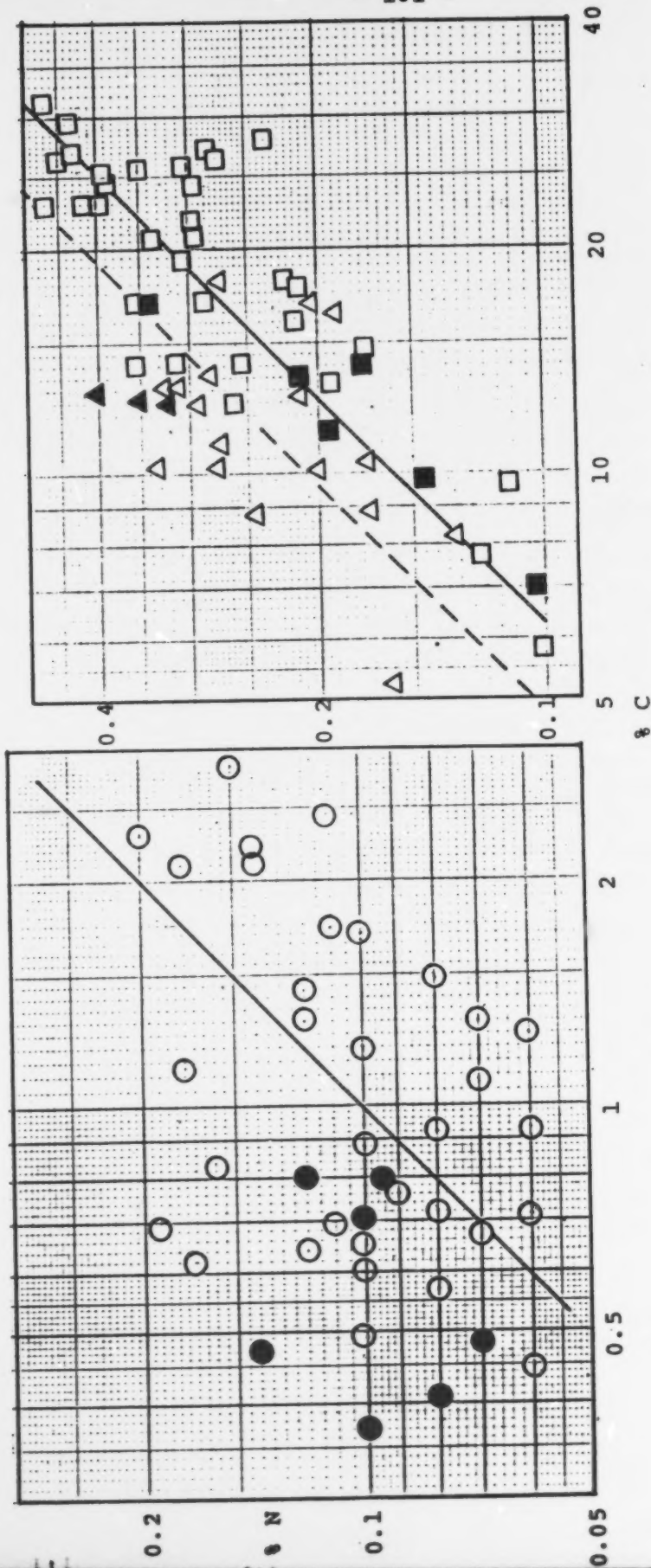


Fig. 10.2. Nitrogen content of solid products vs. the carbon content. Product: $\circ$, bed; $\square$, cyclone; $\triangle$, baghouse. Coals: $\circ$, Minto; $\bullet$, Devco. The full line (bottom graph), $N = 0.016 C$, statistically fits the cyclone data, and the dashed line, $N = 0.021 C$, fits the baghouse data. The full line (top graph) represents $N = 0.1 C$. For the Minto coal, $N = 0.012 C$, and for Devco, $N = 0.017 C$.

Table 10.1. Mass ratio N/C in the cyclone and baghouse products. The values for the coals are 0.012 kg/kg for Minto and 0.017 kg/kg for Devco.

Coal	Product	N/C, kg/kg:	
		Ratio	Std. error
Minto	cyclone	0.016	0.005
Devco	cyclone	0.016	0.002
Minto	baghouse	0.021	0.006

10.4 Conversion of fuel N to NO_x

Figures 10.3 and 10.4 show the NO_x yield factor $\dot{n}_{\text{NO}_x} / \dot{n}_{\text{N},f}$ (moles NO_x produced per mole of fuel nitrogen atoms) as a function of various operating variables: bed temperature, excess air level, recycle fraction, solids product mass flux, and bed product mass flux. The last two (as in the case of carbon, Sect. 8.2.1) serve as reciprocal measures of solids residence time. It is seen that only 10--24 % of the fuel nitrogen actually leaves the combustor as NO_x. There is no apparent correlation of the NO_x yield with bed temperature or recycle. There is a suggestion of dependency on the excess air level and/or the residence time $\tau_b = \dot{m}_b / \dot{m}_b$. If the first is true, as seems most likely, then it would appear that by holding the excess air level very low, at 0--3 %, NO_x emission may be cut in half relative to that at excess air levels above 10 %.

10.5 The missing fuel N

The NO_x emissions evidently account for 9--24 % of the fuel nitrogen. Nitrogen in the char carried by the solid products accounts for no more than 15 %. The remainder is unaccounted for by our analyses, and we can only conclude that it must have been converted to molecular nitrogen, N₂.

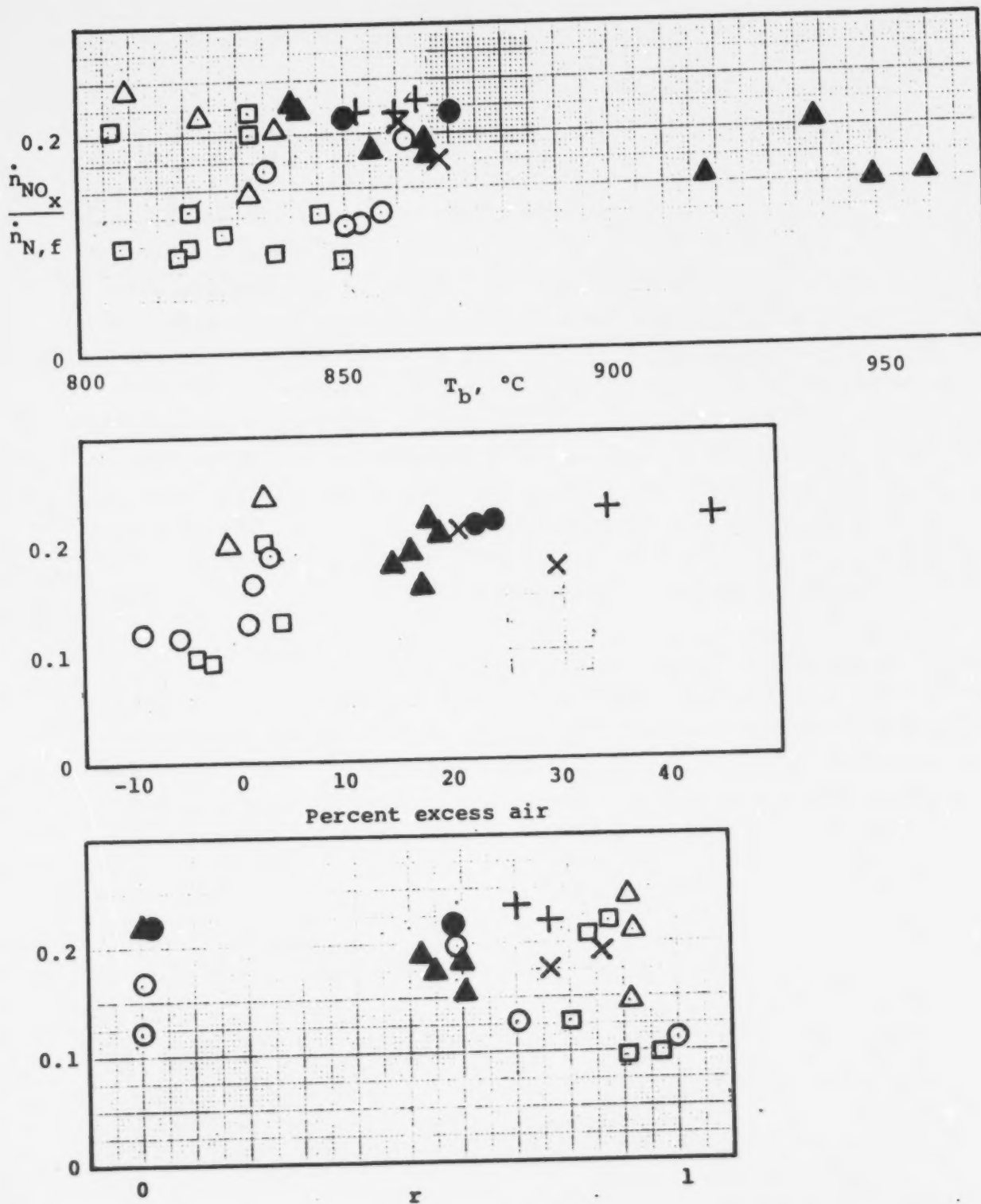


Fig. 10.3. The molar flux ratio $\dot{n}_{NO_x}/\dot{n}_{N,f}$ (the fuel nitrogen to NO_x conversion factor) at combustor exit vs. bed temperature (top), excess air level (centre), and recycle fraction (bottom). Key to data: Table W.3B.

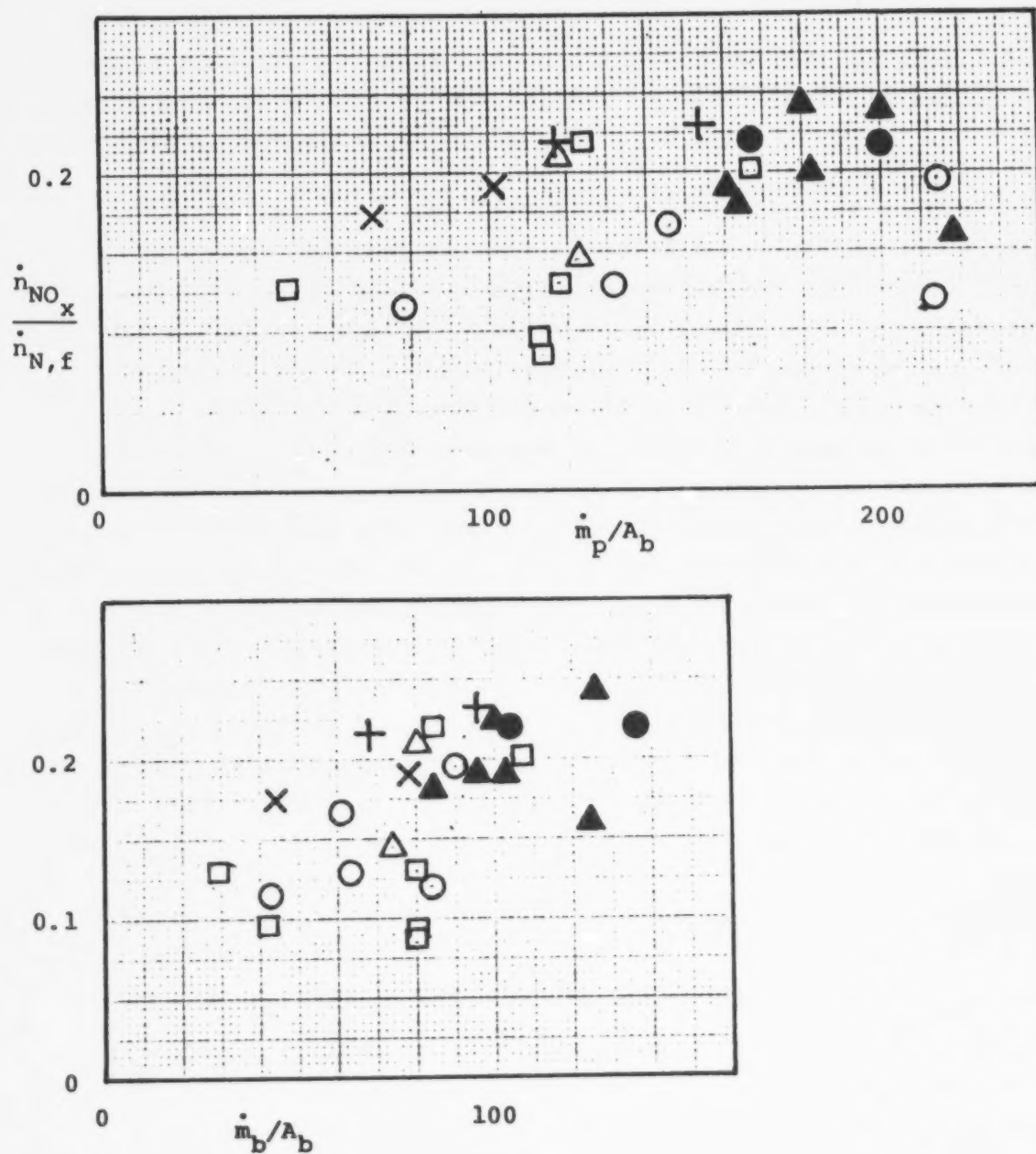


Fig. 10.4. The molar flux ratio $\dot{n}_{NO_x}/\dot{n}_{N,f}$ (the fuel nitrogen to NO_x conversion factor) at combustor exit vs. total solid products flux per unit bed area (top) and bed product flux per unit area (bottom). Key to data: Table N.3B.

11. THE SULPHUR BALANCE

11.1 Introduction

Sulphur enters the system as fuel sulphur. The "fuel" sulphur, of course, is different from the fuel hydrogen and fuel carbon in that it is only partly organic, a major portion being inorganic (pyritic, bound up in metal sulphides, chiefly with iron). Sulphur leaves the system partly in the solid products--as "organically" bound sulphur in the char and as inorganic sulphur in unreacted metal sulphides and, most importantly, as sulphate sulphur in sulphated lime. The remaining portion leaves as SO_2 and SO_3 in the exhaust gases. The SO_3 , however, is negligible. Our analyzer for sulphur in solids does not resolve organic and inorganic sulphur, nor sulphide and sulphate sulphur; we thus simply measured the total sulphur content of the coal and the solid products.

The aim of feeding limestone in fluidized bed combustion, of course, is to capture SO_2 so it will not be emitted to the atmosphere. We have

$$\dot{n}_{\text{SO}_2, x} + \dot{n}_{\text{S}, p} = \dot{n}_{\text{S}, f} , \quad (11.1.1)$$

where $\dot{n}_{\text{SO}_2, x}$ is the molar flux of SO_2 leaving the system in the gaseous products, $\dot{n}_{\text{S}, p}$ is the molar flux of sulphur leaving in the solid products, and $\dot{n}_{\text{S}, f}$ is the molar flux of sulphur entering in the fuel. The nominal SO_2 capture efficiency is simply defined as

$$\eta_{\text{SO}_2} = 100(1 - \dot{n}_{\text{SO}_2, x} / \dot{n}_{\text{S}, f}) = 100 \dot{n}_{\text{S}, p} / \dot{n}_{\text{S}, f} . \quad (11.1.2)$$

It is called nominal because some of the sulphur--normally a quite minor fraction--that does not escape as SO_2 was never oxidized, so its absence in the emissions is not actually indicative of SO_2 capture. This is the sulphur that is bound in

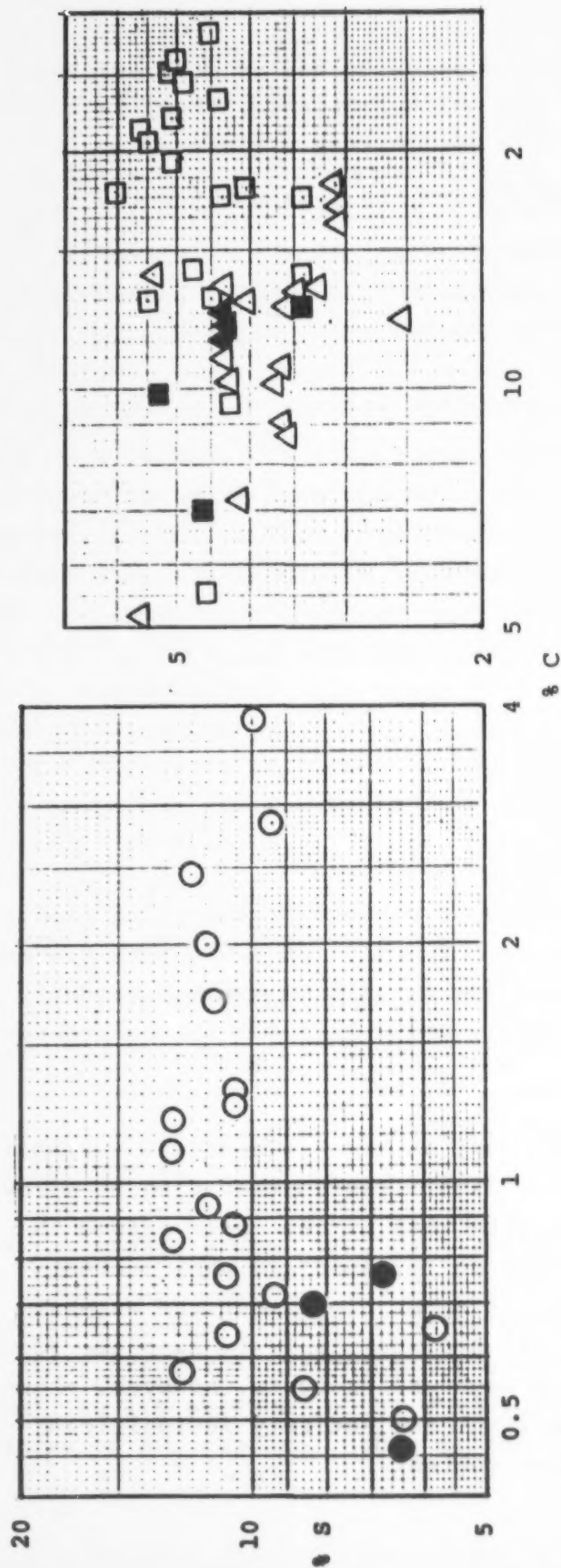
the solid products as unreacted sulphide and as char sulphur. A major concern in examining the sulphur balance, then, is to determine the behaviour of the SO_2 capture efficiency η_{SO_2} . So long as the combustion efficiency is reasonably good, η_{SO_2} essentially indicates the effectiveness of the limestone in SO_2 removal, and the practical objective is to define the conditions under which η_{SO_2} is acceptably high.

The economically most important variable affecting η_{SO_2} is the level of application of limestone. This is commonly expressed by the molar ratio of alkaline earth acceptor atoms in the limestone to sulphur atoms in the coal. Because our limestone was nearly (97 %) pure CaCO_3 , we will call this the Ca/S ratio. We will however include in the "Ca" the small amount of magnesium present in the limestone. We will not take into account the calcium and magnesium present in the coal. For coals with a substantial alkaline earth content, appropriate recognition would, though, be important.

11.2 Sulphur content of the solid products

Figure 11.1 shows the sulphur content of the solid products (bed, cyclone and baghouse) as a function of the carbon content. The ranges of sulphur content are 6--13 % in the bed product, 3.4--6 % in the cyclone product, and 2.5--5.6 % in the baghouse product.

The choice of carbon content as the "independent variable" in Fig. 11.1 is of no special significance other than to allow comparison of the S/C ratio with that of the coal. Since the char as such is likely to have an "organic" sulphur content similar to that of the coal, it seems clear that the sulphur content of the bed product was overwhelmingly in sulphate (CaSO_4) sulphur. In the cyclone and baghouse products, however, organic sulphur may have been a small but not entirely insignificant fraction of the total sulphur content.



11.3 Distribution of sulphur flux in the products

Figure 11.2 shows the distribution of sulphur flux in the products. As expected, the bed product was the major carrier, taking 62--95 % of the total. The baghouse product figured largely only in a run at full recycle when it took 25 % of the total; in the rest, it had only 1--5 % in the Minto trials and 5--7 % in the Devco. The cyclone product took 4--24 % in the runs with partial recycle.

11.4 The SO₂ capture efficiency based on product gas analysis

The SO₂ capture efficiency is expected to depend most strongly on the Ca/S ratio and on the limestone particle size. This is clearly evident in Fig. 11.3 which shows the SO₂ capture efficiency based on the product gas analysis,

$$\eta_{\text{SO}_2} = 100(1 - \dot{n}_{\text{SO}_2, \text{x}} / \dot{n}_{\text{S}, \text{f}}) ,$$

as a function of Ca/S. The data symbols convey additional information about the operating conditions: the mean limestone particle size, the level of fluidizing velocity, the temperature range, and the coal type. There is a suggestion from the Minto trials (with 0.78 mm limestone) that increasing the fluidizing velocity at constant Ca/S tends to decrease η_{SO_2} , but only weakly. Since substantial increase in the fluidizing velocity was generally accompanied by proportional changes in coal and (at constant Ca/S) limestone feedrates, this effect is likely largely due to decreasing solids residence time. The results of the Devco trials (with 1.72 mm limestone), however, indicate no effect of fluidizing velocity.

As expected from the known phenomenon of pore-plugging of limestone during sulphation, which tends to limit effectiveness to an outer layer less than 1 mm deep, increasing limestone

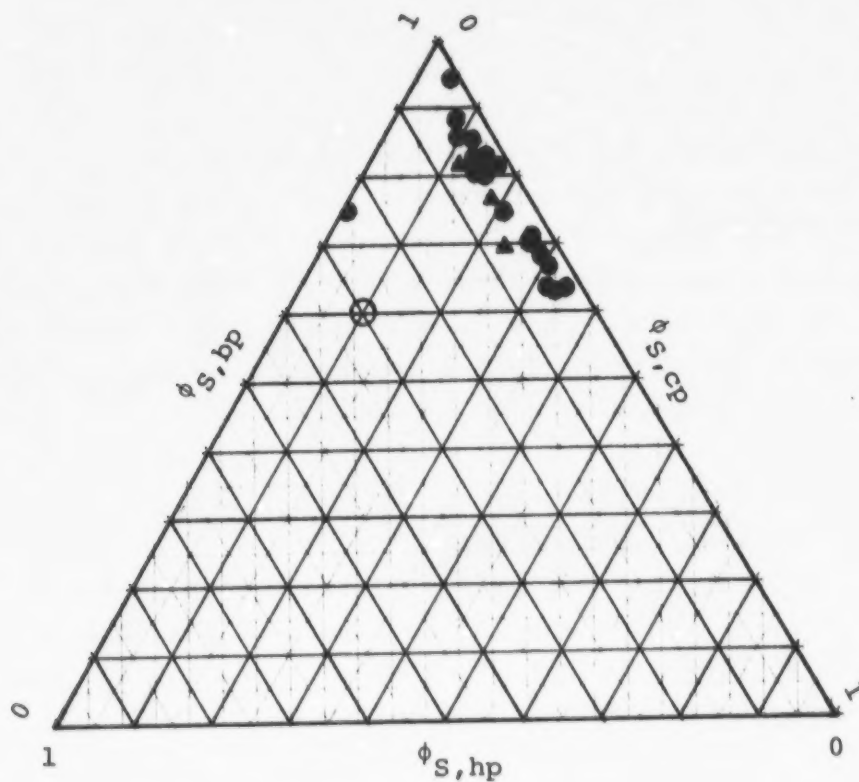


Fig. 11.2. Partition of sulphur flux between bed product, cyclone product, and baghouse product. Coals: ●, Minto; ▲, Devco. Illustrative point: $\phi_{S,bp} = 0.6$, $\phi_{S,cp} = 0.1$, $\phi_{S,hp} = 0.3$.

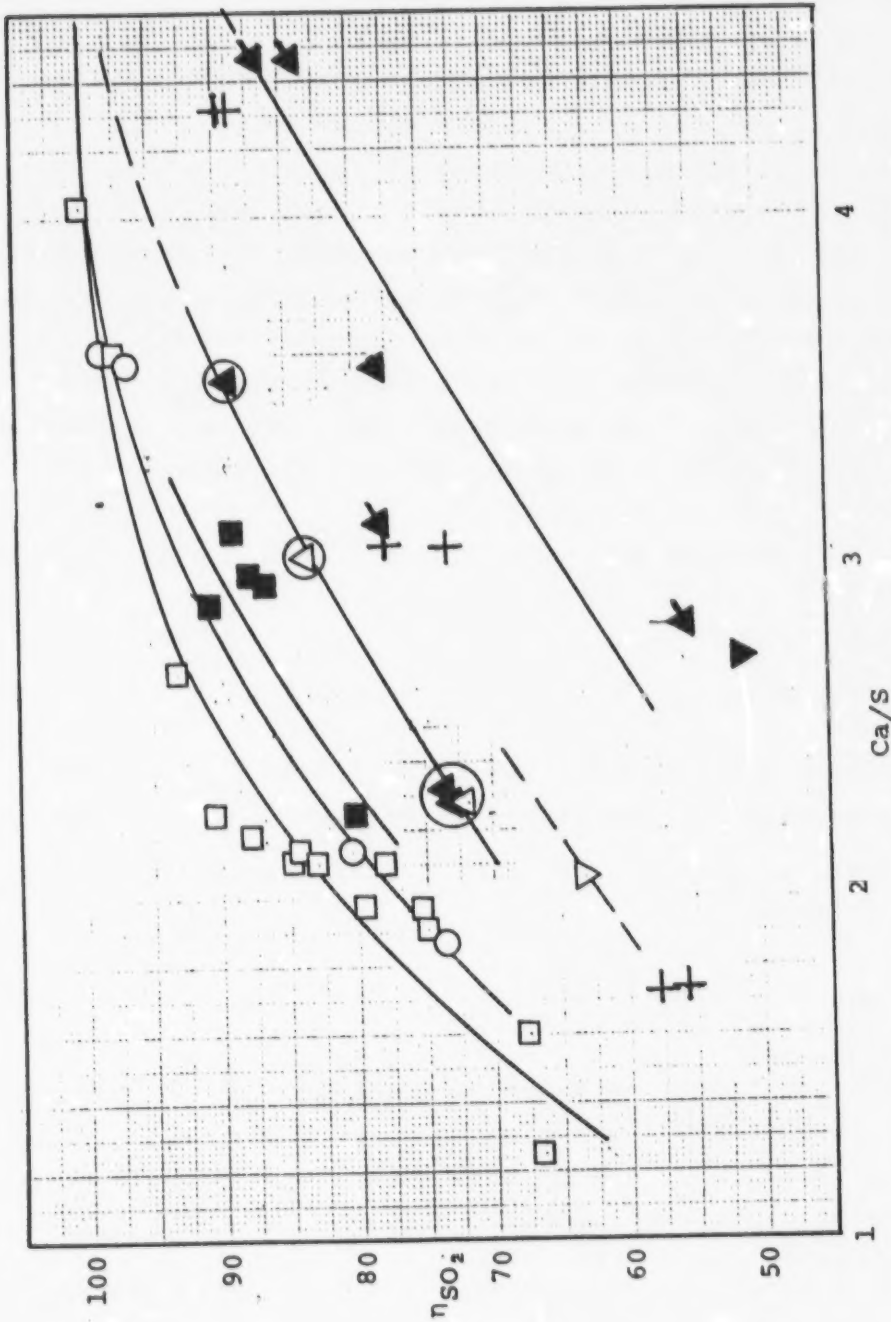


Fig. 11.3. Percent sulphur capture vs. Ca/S ratio, capture based on gas analysis at combustor exit. Data source: +, Stover, et al. (1981); other data, present work. Limestone particle diameter $\hat{D}_p$, mm: $\bigcirc$, 0.49; $\square$, 0.78; $\triangle$, 1.72; ∇ , 1.98; +, 1.15. Coals: circled points, Devco; other, Minto. Levels of fluidizing velocity: $\bigcirc \square \triangle \nabla$, 1.6--2.0; +, 2.2; $\blacksquare \blacktriangle \blacktriangledown$, 2.7--3.0. Flagged points ($\blacktriangle$): $T_b > 900^\circ\text{C}$ (918--959 $^\circ\text{C}$).

particle size much over 1 mm dia strongly diminishes SO₂ capture efficiency. On the other hand, decreasing particle size much under 1 mm improves it little or negligibly.

Figure 11.4 shows a resolution of the data on the 0.49 mm and 0.78 mm dia limestones on the basis of bed temperature. Comparing Figs. 11.3 and 11.4 indicates no apparent effect of temperature over the experimental range of these results, 800--870 °C. The data are evidently best resolved on the basis of limestone particle diameter and fluidizing velocity alone, as in Fig. 11.3. The curve fitted to the data for 840--859 °C in Fig. 11.4 does, though, provide a reasonable mean representation of the results on the small limestone particles. The Ca/S ratios on this curve at various interesting SO₂ capture efficiencies are:

η_{SO_2}	80	90	95	98	99
Ca/S:	2.1	2.8	3.2	3.6	3.9

Figure 11.5 shows the data from Fig. 11.3 on a different graph--a semilogarithmic representation of $(100 - \eta_{SO_2})$ vs Ca/S. Within the experimental range of Ca/S, 1.2--4.4 a straight-line fit is good, giving

$$\eta_{SO_2} = 100 - A e^{-B(Ca/S - 1)} \quad (11.4.1)$$

The constants A and B for the curves drawn are given in the caption under the figure.

11.5 The sulphur capture efficiency based on product solids analysis

Figure 11.6 shows as a function of the Ca/S ratio the SO₂ capture efficiency based on the product solids analysis,

$$\eta_{SO_2} = 100 \dot{n}_{S,p} / \dot{n}_{S,f} \quad ,$$

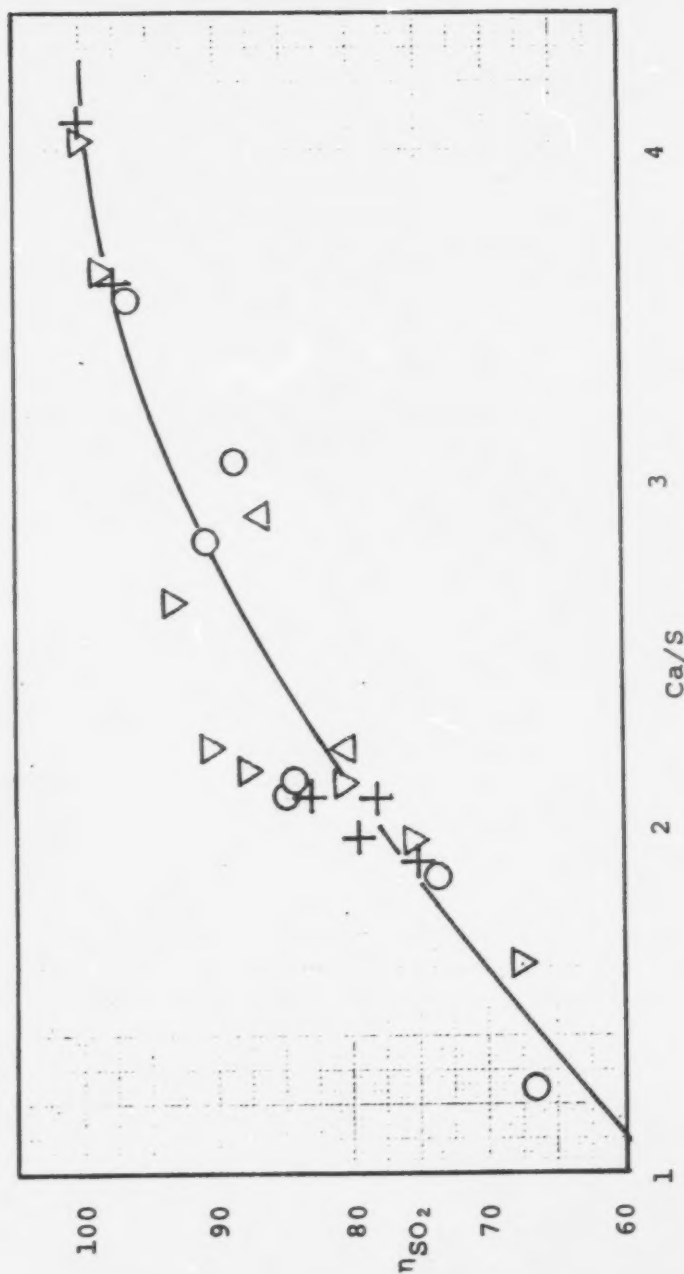


Fig. 11.4. Percent sulphur capture vs. Ca/S ratio, capture based on gas analysis at combustor exit. Data for limestone particle diameters D_p of 0.49 mm and 0.78 mm. Bed temperature ranges, °C: +, 800--819; ▽, 820--839; O, 840--859; Δ, 860--880.

Fig. 11.5. Percent sulphur escaping in SO₂ emissions vs. Ca/S.
See Fig. 11.1 for key to data. The straight lines represent the equation

$$\eta_{\text{SO}_2} = 100(1 - A e^{-B(\text{Ca/S} - 1)}) ,$$

where η_{SO_2} is the sulphur capture efficiency. Values of A and B are:

Curve	a	b	c	d	e
A	0.70	0.85	0.70	0.70	0.70
B	1.33	1.33	0.94	0.73	0.45

See figure on next page

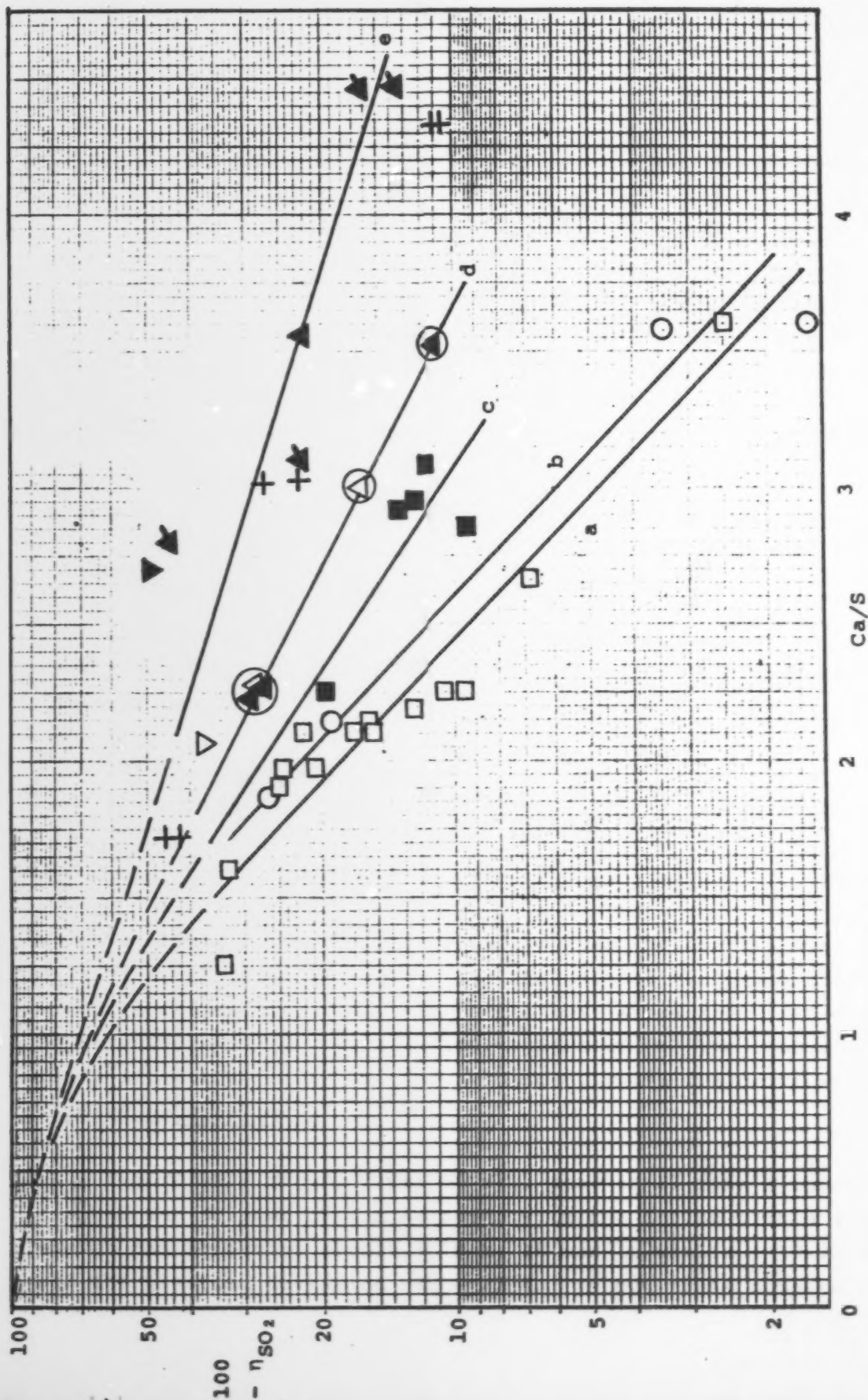


Fig. 11.5. See caption on preceding page.

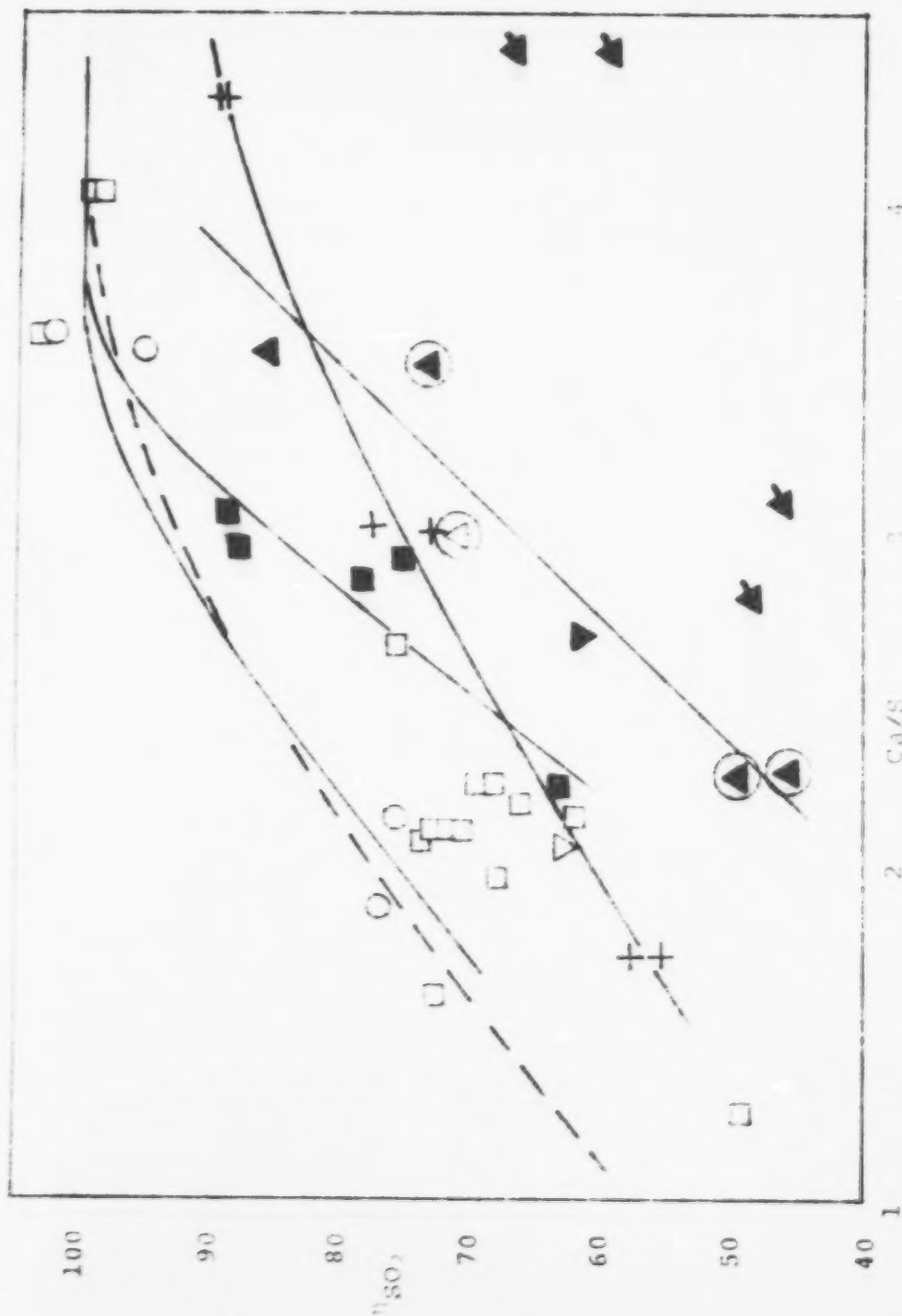


Fig. 11.6. Percent sulphur capture vs. Ca/S ratio, based on solids analysis. See Fig. 11.1 for key to data. Dashed curve: the curve from Fig. 11.4.

$$\begin{aligned}\dot{n}_{S,p} &= \dot{n}_{S,b} + \dot{n}_{S,c} + \dot{n}_{S,h} \\ &= (\dot{m}_b w_{S,b} + \dot{m}_c w_{S,c} + \dot{m}_h w_{S,h}) / A_S\end{aligned}$$

where A_S is the molar atomic mass of sulphur. If the system sulphur balance were perfect, this graph should be identical with Fig. 11.3, the corresponding graph based on gas analysis. The results for the smallest limestone particle size, 0.49 mm mean dia, do indeed agree well, and so do those for the 0.78 mm particles at large Ca/S (when η_{SO_2} is about unity). The other results, however, seem to show significant discrepancies, generally with the solids analysis indicating lower values of η_{SO_2} . This is clearly seen in Figs. 11.7 and 11.8, graphs of η_{SO_2} as indicated by the solids analysis vs. η_{SO_2} as indicated by the gas analysis. The problem of the discrepancies basically revolves around the total sulphur balance which is discussed in the next section.

11.6 The total sulphur balance

The steady-state sulphur balance, (11.1.1), requires that

$$(\dot{n}_{SO_2,x} + \dot{n}_{S,p}) / \dot{n}_{S,f} = 1 \quad (11.6.1)$$

We have examined many hypotheses as to what the experimental deviations of this flux ratio from the theoretical value of unity might be related to. The only factor that shows any visible degree of correlation, Fig. 11.9, is the temperature at the combustor exit, after the freeboard heat exchanger, where the critical gas samples were taken.

Figure 11.9 shows, first of all, that in 21 runs a total of six had the sulphur efflux agreeing with the influx within $\pm 5\%$, eleven had agreement within $\pm 10\%$, 18 had it within $\pm 10-20\%$, and the greatest deviation was -27% . To put these numbers in

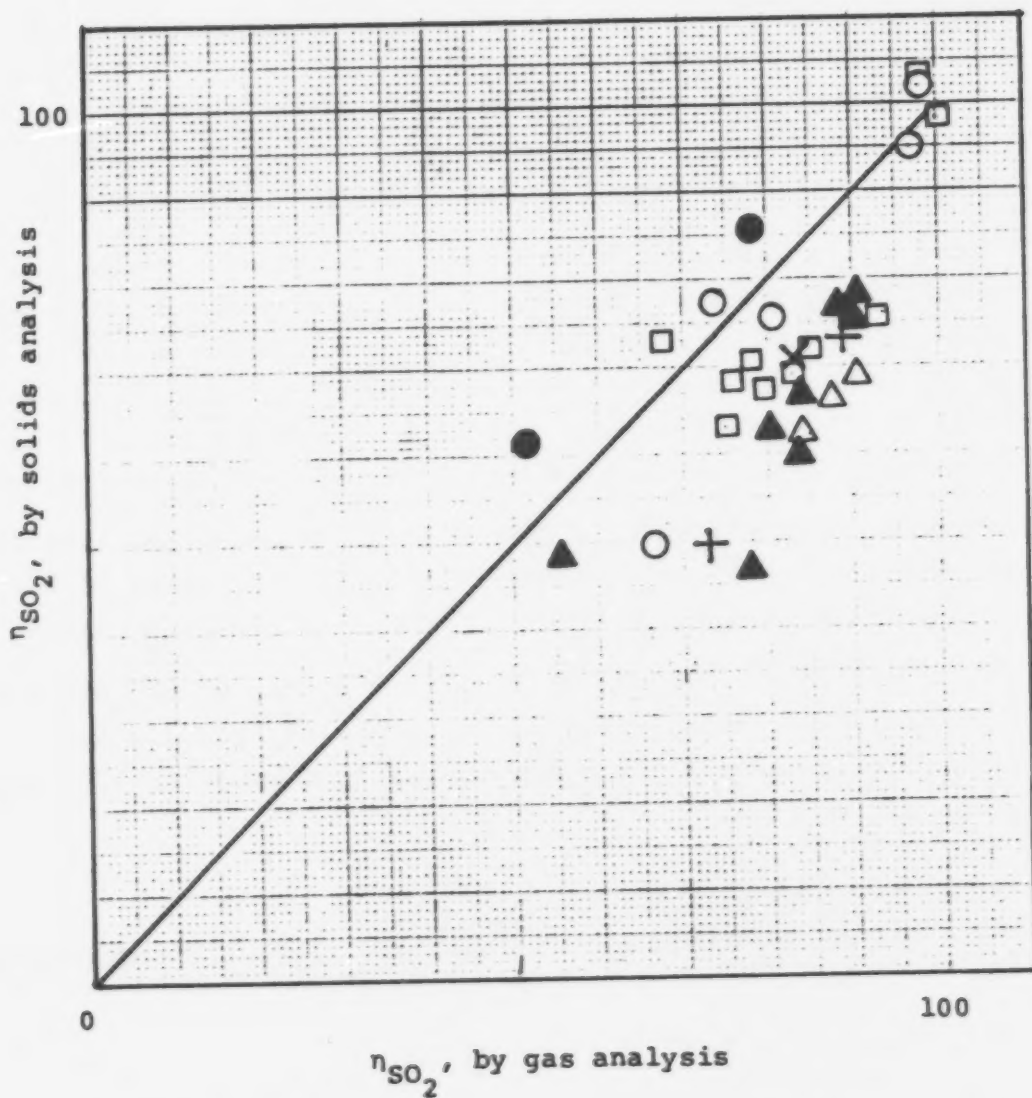


Fig. 11.7. Sulphur capture according to solids analysis vs. that based on gas analysis at combustor exit. Key to data: Table N.3B.

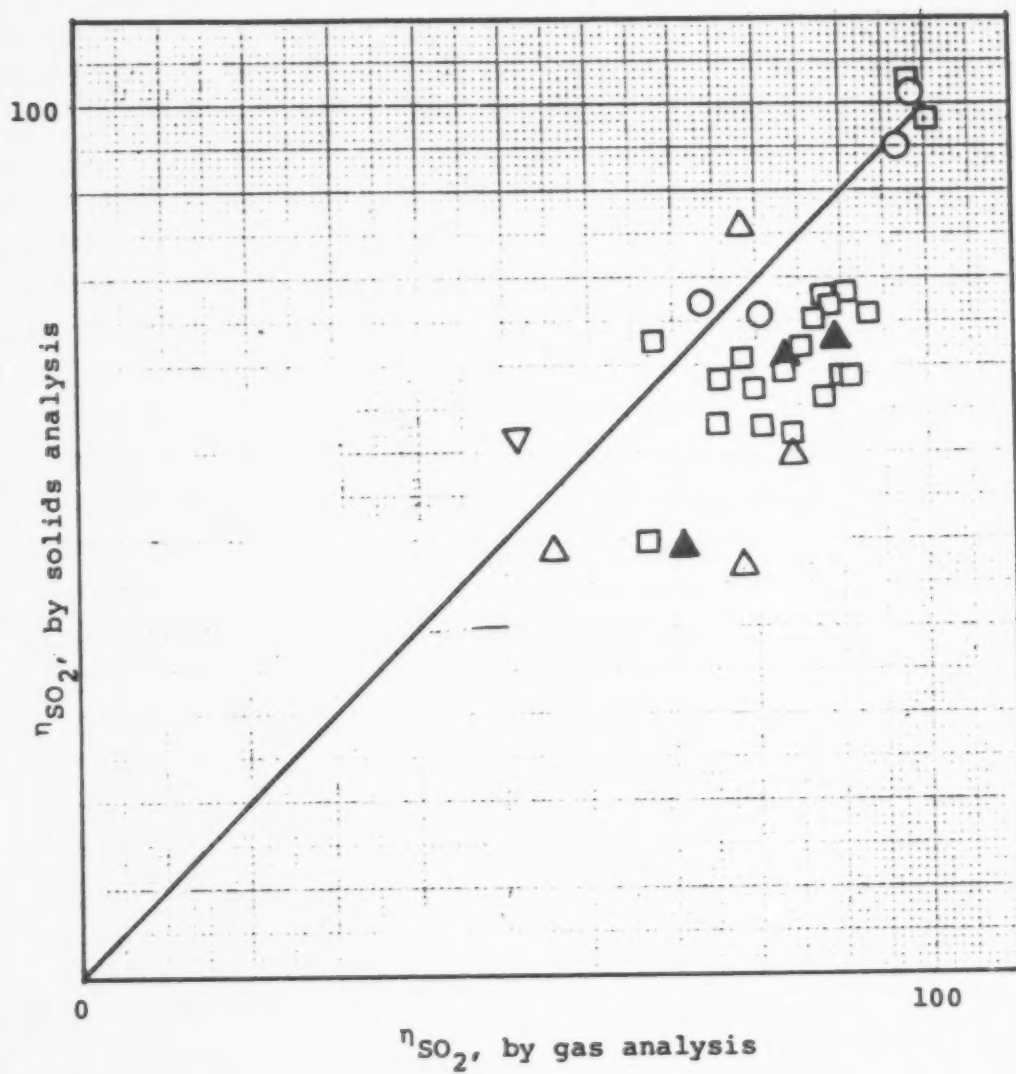


Fig. 11.8. Sulphur capture according to solids analysis vs. that based on gas analysis at combustor exit. Key to data: Table N.3A.

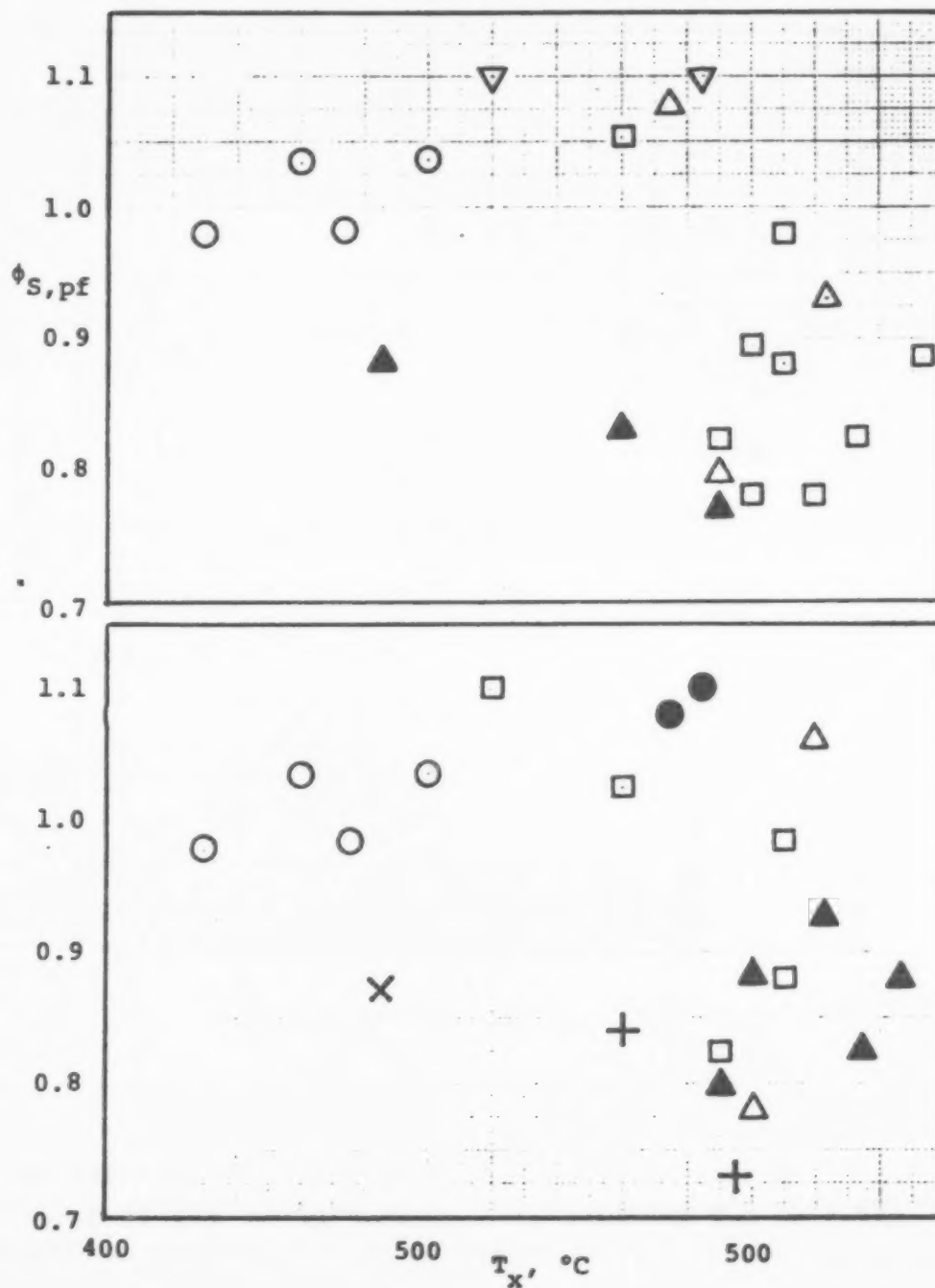


Fig. 11.9. Flux ratio of sulphur in all products to sulphur fed in fuel, vs. temperature at bed exit. Top, limestone identified (key in Table N.3A); bottom, coal identified (key in Table N.3B).

perspective, consider a run in which the sulphur balance was not very good, Run 830309 with Devco coal (the datum denoted by the symbol x in Fig. 11.9). The run data indicate that the coal feedrate was 29.1 ± 1.7 kg/hr, the bed product efflux was 14.4 ± 0.9 kg/hr, the cyclone product efflux was 2.85 ± 0.38 kg/hr, and the baghouse product efflux was 1.6 ± 0.2 kg/hr. The sulphur analyses were 5.17 % for the coal (as fed--i.e., wet basis) and, respectively, 8.34 %, 4.58 % and 4.34 % for the bed, cyclone and baghouse products. Suppose, for this argument, that the sulphur analyses were perfect (the precision was ± 0.13 % on the coal, for example, but we do not have data for a rigorous assessment of accuracy). The value of the efflux/influx sulphur ratio is then, in symbols,

$$\begin{aligned} & (\bar{Y}_1 + y_1 + \bar{Y}_2 + y_2 + \bar{Y}_3 + y_3) / (\bar{X} + x) \\ & = ((\bar{Y}_1 + \bar{Y}_2 + \bar{Y}_3) / \bar{X} + \eta_1 + \eta_2 + \eta_3) (1 - \xi) \\ & = \bar{E} - \bar{E} \xi + \eta - \eta \xi \end{aligned}$$

where

$$y_1 = \dot{m}_b w_{S,b} , \quad y_2 = \dot{m}_c w_{S,c} ,$$

$$y_3 = \dot{m}_h w_{S,h} , \quad x = \dot{m}_f w_{S,f} ,$$

$$Y_1 = \bar{Y}_1 + y_1 , \quad Y_2 = \bar{Y}_2 + y_2 ,$$

$$Y_3 = \bar{Y}_3 + y_3 , \quad X = \bar{X} + x ,$$

$$\bar{E} = (\bar{Y}_1 + \bar{Y}_2 + \bar{Y}_3) / \bar{X} ,$$

$$\eta_1 = y_1 / \bar{X} , \quad \eta_2 = y_2 / \bar{X} ,$$

$$\eta_3 = y_3 / \bar{X} , \quad \xi = x / \bar{X} ,$$

$$\eta = \eta_1 + \eta_2 + \eta_3 .$$

The mean square deviation in this ratio is thus

$$\begin{aligned} s^2 &= \overline{(-\bar{E} \xi + \eta - \eta \xi)^2} \\ &= \bar{E}^2 \overline{\xi^2} + \overline{\eta^2} + \overline{\eta^2 \xi^2} - 2 \bar{E} \overline{\xi \eta} \\ &\quad - 2 \bar{E} \overline{\xi \eta} - 2 \overline{\eta^2 \xi} \end{aligned}$$

Assuming no correlations between ξ , η_1 , η_2 , and η_3 , this reduces to

$$s^2 = \bar{E}^2 \overline{\xi^2} + \overline{\eta^2}$$

$$\overline{\eta^2} = \overline{\eta_1^2} + \overline{\eta_2^2} + \overline{\eta_3^2} .$$

Now

$$\begin{aligned} \bar{E} &= ((14.4)(0.0834) + (2.85)(0.0458) \\ &\quad + (1.6)(0.0434)) / (29.1)(0.0517) \\ &= 0.931 \end{aligned}$$

$$\overline{\xi^2} = (1.7/29.1)^2 = 0.00341$$

$$\begin{aligned} \overline{\eta^2} &= \{[(0.9)(0.0834)]^2 + [(0.38)(0.0458)]^2 \\ &\quad + [(0.2)(0.0434)]^2\} / [(29.1)(0.0517)]^2 \\ &= 0.00266 \end{aligned}$$

Thus

$$s^2 = (0.931)^2(0.00341) + 0.00266 = 0.00562$$

$$s = 0.075$$

Our conclusion is that the standard error in the efflux/influx sulphur ratio due just to the uncertainties in the coal and solid product mass fluxes in Run 830309 is $\pm 7.5\%$.

The uncertainty in the solid fluxes was significantly lower in some runs than in this example, and in some it was a little higher. It appears, in short, that the standard error in the efflux/influx sulphur ratio under optimum operating conditions was around $\pm 4\%$, and under adverse conditions it was typically around $\pm 8\%$. Thus deviations of up to $\pm 10\%$ in Fig. 11.8 are well within the expected norm. These observations are in reasonable concordance with the carbon balance--which was generally good--as seen in Fig. 8.15. Deviations from unity in the efflux/influx carbon ratio all fell within a $\pm 8\%$ band.

The large negative deviations from unity of some of the efflux/influx sulphur ratios in Fig. 11.9, though, and certainly those exceeding -15% , likely reflect some measurement error or operating problem which we have not yet been able to identify conclusively. The simple inference from Fig. 11.9 is that a significant amount of SO_2 was being formed, presumably by char combustion or by breakdown of CaSO_4 , beyond the combustor-exit gas sampling point and while the gases were above 580°C . However, this view of the matter may be naive, and it is possible that Fig. 11.9 is entirely misleading, based on a fortuitously prejudiced set of operating conditions.

In planning our experiments, we assumed that chemical reaction would be quenched by the overbed heat exchanger, that reactions of char and CaSO_4 would be negligible in the time available below 650°C . The combustor exit was fitted with a very convenient gas sampling port and probe, and no measurements

were made further downstream. Now it appears that we should have checked SO_2 levels at several points downstream (say after the cyclone and after the baghouse). Such measurements will be made in future work. As already noted, the freeboard heat exchanger will also be removed and replaced entirely by an efficient outboard heat exchanger. The point of quenching will then be quite sharply defined, and the quenching process should be first and complete.

11.7 Solids sulphur analyses at intermediate points

Samples of bed material collected with a probe directly from the bed at ports 244 mm and 569 mm above the distributor plate as described in Sect. 8.5 were analyzed for sulphur. Figure 11.10 shows that the sulphur contents probably differed negligibly from those of the bed product which was collected by overflow from the top of the bed. In terms of sulphur content, then, the bed appears to have been well mixed.

Samples of elutriating particles collected with a dust sampling probe from the gases at bed exit agreed quite closely in sulphur content, Fig. 11.11, with the values computed from the analyses of the cyclone and baghouse products.

We see in these results only consistency, with no clues to an explanation of discrepancies in the overall sulphur balance.

11.8 Closure

The results show that Havelock limestone is a good limestone for SO_2 capture in fluidized bed combustion; i.e., good SO_2 capture can be had at reasonable Ca/S ratios. It is important, however, that the particle size not be too large. Material with a mean particle diameter of 0.5--1 mm seems to be optimum. Particles much larger than this require considerably higher Ca/S ratios for a given capture efficiency. Except when the fluidizing velocity is set uninterestingly low (diminishing system capacity), particles much smaller than this tend to be quickly

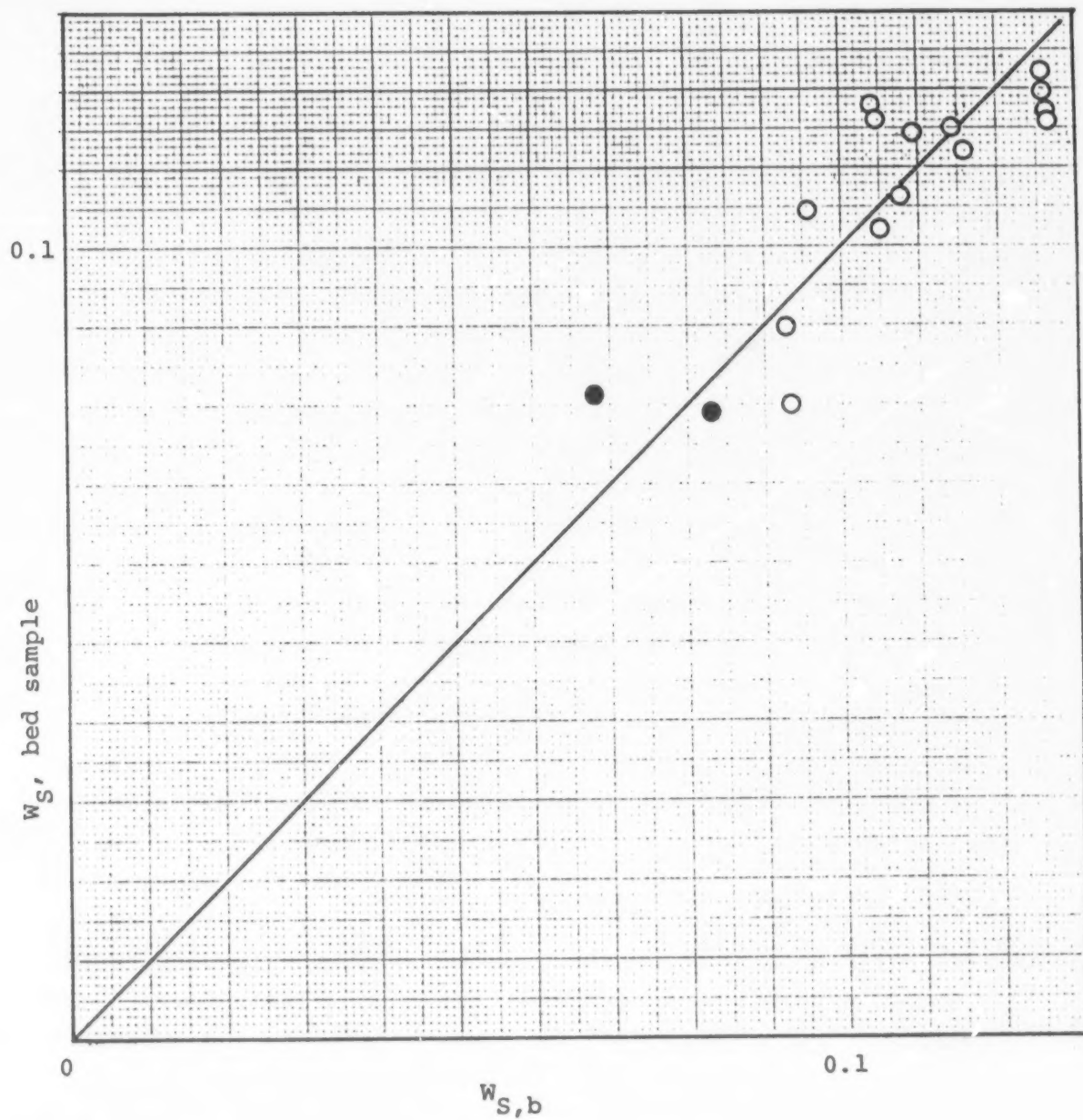


Fig. 11.10. Sulphur in solids sample from bed vs. that in the bed product. Bed samples from ports 3 and 34, Fig. 14.2. Coals: ○, Minto; ●, Devco.

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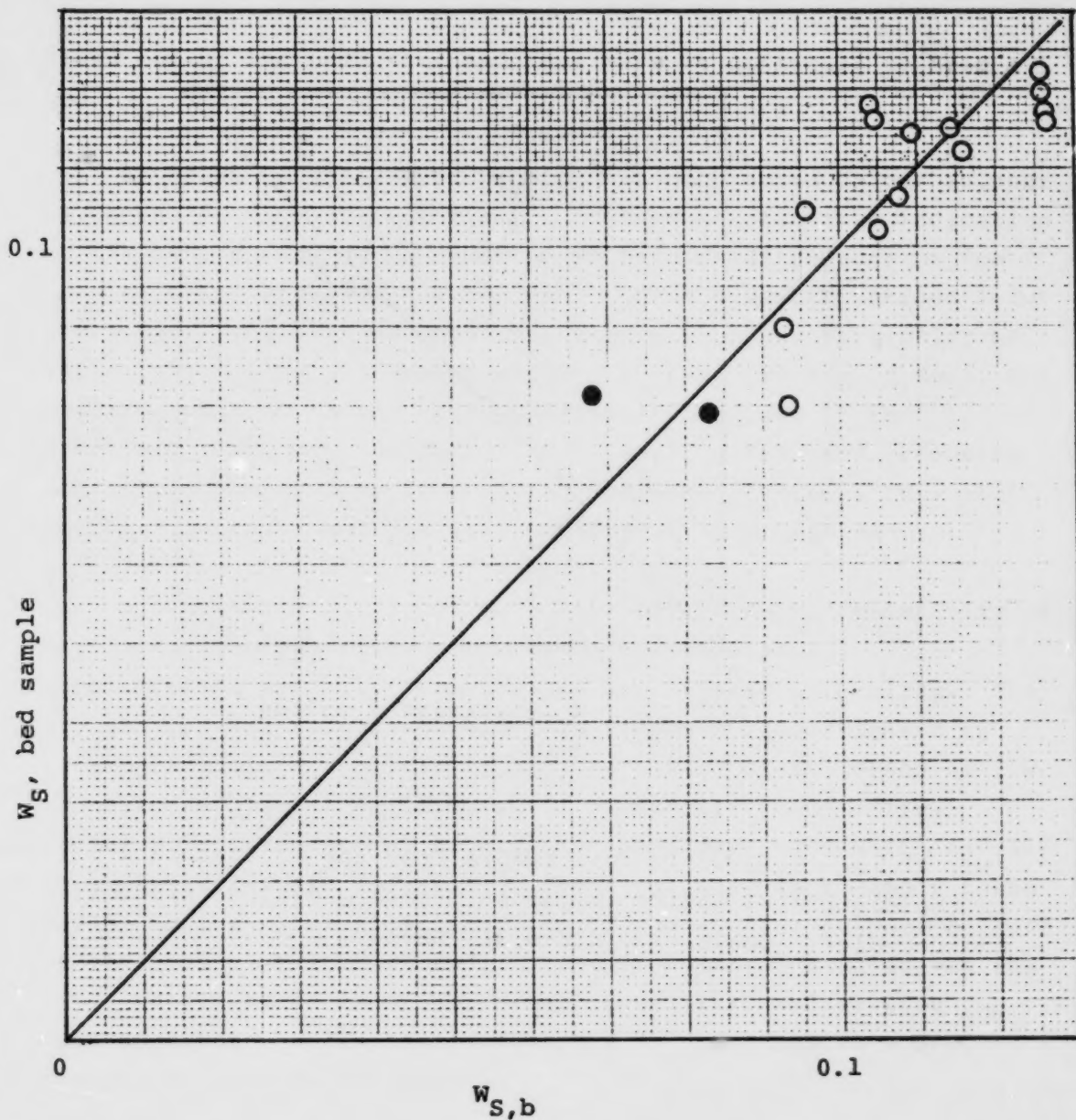


Fig. 11.10. Sulphur in solids sample from bed vs. that in the bed product. Bed samples from ports 3 and 34, Fig. 14.2. Coals: $\circ$, Minto; $\bullet$, Devco.

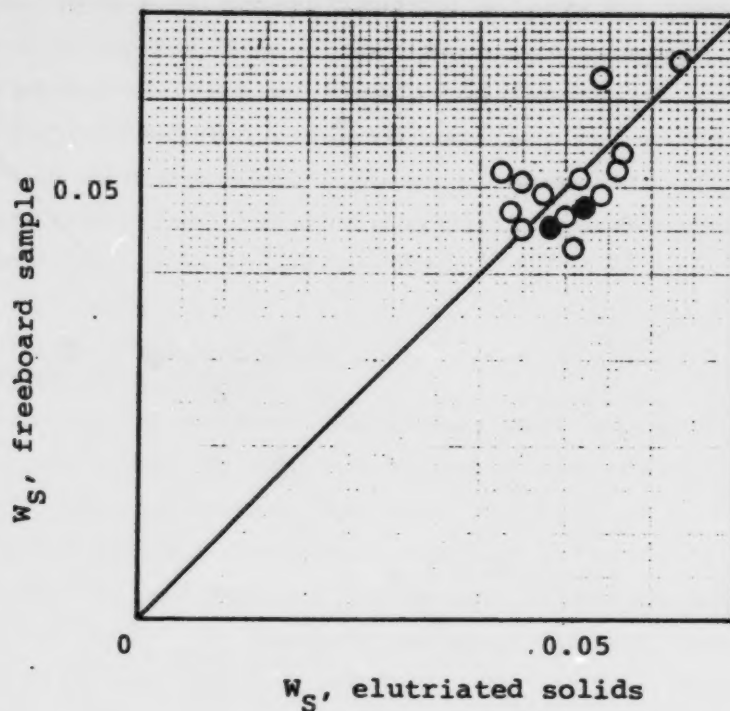


Fig. 11.11. Sulphur in dust sample from top of freeboard vs. that in the elutriated solids (baghouse product, cyclone product, and recycle). Coals: ○, Minto; ●, Devco.

blown out of the combustor and thus make it difficult to hold a bed.

Although an unresolved problem has been noted in our sulphur balances, it appears that Fig. 11.4 gives a reasonable indication of the performance to be expected of Havelock limestone in the mean particle size range 0.5--1 mm. Certain qualifications about the general validity of results obtained with any particular FBC installation should, however, be remembered. We saw in Sect. 6.4 that SO_2 content of the gases sometimes increased in the freeboard and sometimes decreased. Clearly freeboard processes add an element of apparatus dependence; the question, "How high is the freeboard, and how is it shaped?", is significant.

A fundamental study of the characteristics and reactivity of Havelock limestone is in progress in our laboratory, complementing the pilot plant work, and is nearly complete. The calcination and sulphation kinetics are being investigated by thermogravimetric analysis (TGA). Pore size distribution of raw and calcined limestone is measured by mercury porosimetry. Mechanistic mathematical models have been developed which explain the kinetic results very well. The final task is to apply these models to the pilot plant results, to see whether sulphur capture is reasonably well predicted. These fundamental studies are the Ph.D. thesis research of Mr. Michel Couturier, and will be published through his thesis (which will be issued as a QFBC report) and in journal papers.

12. OTHER ELEMENT BALANCES: Ca, Mg, Na, K, Fe, Si, V

12.1 Introduction

Analyses for the major, and for certain significant minor, elemental constituents of ash and limestone were obtained for four runs. The analyses were performed by a commercial analyst, and the number of the analyses was limited because of cost. Some of the major elements, particularly iron and silicon, occurred very predominantly in the coal and serve as tracers for the bulk of the coal ash. Others, namely calcium and magnesium, occurred very predominantly in the limestone.

12.2 The results

The results are summarized in Tables 12.1--12.4 which show the element mass fraction in each solids stream, $w_{i,m}$; the mass flux of that element carried by the stream, $\dot{m}_{i,m} = \dot{m}_m w_{i,m}$; and the fraction of the total element influx or efflux carried by the stream, $\phi_{i,mn} = \dot{m}_{i,m} / \dot{m}_{i,n} = \dot{m}_{i,m} / \sum \dot{m}_{i,m}$, where the summation is either over all the feed streams or all the product streams.

12.3 The overall element balance

In steady-state operation, the total efflux of an element in all the product streams should equal the influx in all the feed streams. Figure 12.1 shows the experimental efflux/influx ratios for the various elements that were analyzed for, including two that have already been discussed--carbon and sulphur. Among the major elements, the iron and silicon balances (with the exception of one datum for iron) are fairly good. The magnesium, sodium, potassium and vanadium balances are consistent with a relatively high uncertainty due to the small quantities of these elements present. The results for calcium and aluminum, however, are poor, the apparent effluxes being significantly smaller than the influxes. In the case of calcium, the reported calcium content of the limestone in Tables 12.1--12.3, 38.8 %, agrees reasonably well with an independent determination of 39.52 % on stone from

Table 12.1. Elemental analyses and element balances, Run 830201. Operating conditions: Minto coal, 6.6 mm; Havelock limestone, 0.78 mm; excess air = - 1.5 %; Ca/S = 2.18; r = 0.919; v = 1.86 m/s; T_b = 830 °C.

s	m _g kg/hr	Element: Si			Element: Al			Element: Fe			Element: V		
		w _i %	m _{i,s} kg/hr	φ _{i,s}	w _i %	m _{i,s} kg/hr	φ _{i,s}	w _i %	m _{i,s} kg/hr	φ _{i,s}	w _i ppm	m _{i,s} kg/hr	φ _{i,s}
Coal	33.3	2.65	0.882	0.920	0.75	0.250	0.740	4.23	1.409	0.956	11	0.00037	0.66
Stone	17.5	0.44	0.077	0.080	0.50	0.088	0.260	0.37	0.065	0.044	11	0.00019	0.34
I Feed	--	--	0.959	1.00	--	0.338	1.000	--	1.474	1.000	--	0.00056	1.000
Bed	12.4	6.12	0.759	0.679	0.91	0.113	0.351	1.51	0.187	0.187	24	0.00030	0.61
Cycl.	3.6	5.00	0.180	0.161	0.87	0.031	0.096	9.10	0.328	0.328	30	0.00010	0.20
Bagh.	2.2	8.10	0.178	0.160	2.68	0.178	0.552	22.08	0.486	0.485	41	0.00009	0.19
I Prod.	--	--	1.12	1.000	--	0.322	--	--	1.001	1.000	--	0.00049	1.000
		Element: Ca			Element: Mg			Element: Na			Element: K		
		w _i %	m _{i,s} kg/hr	φ _{i,s}	w _i %	m _{i,s} kg/hr	φ _{i,s}	w _i %	m _{i,s} kg/hr	φ _{i,s}	w _i ppm	m _{i,s} kg/hr	φ _{i,s}
Coal	33.3	0.43	0.143	0.021	0.005	0.0017	0.042	0.02	0.007	0.39	0.02	0.007	0.16
Minto	--	--	--	--	--	--	--	--	--	--	--	--	--
Stone	17.5	38.8	6.79	0.979	0.22	0.039	0.958	0.06	0.011	0.61	0.21	0.037	0.84
I Feed	--	--	6.93	1.000	--	0.041	1.000	--	0.018	1.00	--	0.044	1.00
Bed	12.4	30.6	3.79	0.906	0.42	0.0521	0.903	0.08	0.0099	0.66	0.05	0.0062	0.45
Cycl.	3.6	10.52	0.379	0.091	0.13	0.0047	0.081	0.06	0.0022	0.15	0.06	0.0022	0.16
Bagh.	2.2	0.71	0.016	0.004	0.04	0.0009	0.016	0.13	0.0029	0.19	0.25	0.0055	0.39
I Prod.	--	--	4.19	1.000	--	0.0577	1.000	--	0.0150	1.00	--	0.0139	1.00

Table 12.2. Elemental analyses and element balances, Run 830225. Operating conditions: Minto coal, 6.6 mm; Havelock limestone, 1.72 mm; excess air = 19.3 %; Ca/S = 3.10; r = 0.53; v = 2.98 m/s; $T_b = 943^\circ\text{C}$.

	$\dot{m}_g$	Element: Si			Element: Al			Element: Fe			Element: Ca		
		w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$
Coal	39.0	2.65	1.033	0.899	0.75	0.293	0.69	4.23	1.650	0.944	0.43	0.168	0.016
Stone	26.3	0.44	0.116	0.101	0.50	0.132	0.31	0.37	0.097	0.056	38.8	10.20	0.984
I Feed	--	--	1.149	1.000	--	0.425	1.000	--	1.747	1.000	--	10.37	1.000
Bed	14.6	2.62	0.382	0.390	0.72	0.105	0.392	2.44	0.356	0.244	46.97	6.86	0.833
Cycl.	8.5	5.57	0.473	0.483	1.38	0.117	0.436	10.38	0.882	0.604	15.48	1.32	0.160
Begh.	1.32	9.52	0.125	0.127	3.48	0.046	0.172	16.86	0.222	0.152	4.64	0.061	0.007
I Prod.	--	--	0.980	1.000	--	0.268	1.000	--	1.460	1.000	--	8.24	1.00

Table 12.3. Elemental analyses and element balances, Run 830226. Operating conditions: Minto coal, 6.6 mm; Havelock limestone, 1.72 mm; excess air = 17.2 %; Ca/S = 4.47; $r = 0.61$; $v = 293$ m/s; $T_b = 927$ °C

	$\dot{m}_s$ kg/hr	Element: Si			Element: Al			Element: Fe			Element: Ca		
		w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %
Coal	39.4	2.65	1.044	0.851	0.75	0.296	0.59	4.23	1.666	0.916	0.43	0.169	0.010
Stone	41.4	0.44	0.182	0.149	0.50	0.207	0.41	0.37	0.153	0.084	38.8	16.06	0.990
F Feed	--	--	1.226	1.000	--	0.503	1.00	--	1.819	1.000	--	16.23	1.000
Bed	19.2	2.01	0.386	0.346	0.35	0.067	0.295	1.71	0.328	0.201	31.4	6.03	0.600
Cycl.	13.0	4.70	0.611	0.547	0.95	0.124	0.546	8.35	1.086	0.667	20.5	2.66	0.300
Bagh.	1.6	7.51	0.120	0.107	2.25	0.036	0.159	13.44	0.215	0.132	10.8	0.173	0.019
F Prod.	--	--	1.117	1.000	--	0.227	1.000	--	1.629	1.000	--	8.86	1.000

Table 12.2. Elemental analyses and element balances, Run 830225. Operating conditions: Minto coal, 6.6 mm; Havelock limestone, 1.72 mm; excess air = 19.3 %; Ca/S = 3.10; r = 0.53; v = 2.98 m/s; $T_b = 943^\circ\text{C}$.

	Element: Si			Element: Al			Element: Fe			Element: Ca		
	$\dot{m}_s$	w_i	$\dot{m}_{i,s}$	$\phi_{i,s}$	w_i	$\dot{m}_{i,s}$	$\phi_{i,s}$	w_i	$\dot{m}_{i,s}$	w_i	$\dot{m}_{i,s}$	$\phi_{i,s}$
		%	kg/hr		%	kg/hr		%	kg/hr	%	kg/hr	
Coal	39.0	2.65	1.033	0.899	0.75	0.293	0.69	4.23	1.650	0.43	0.168	0.016
Stone	26.3	0.44	0.116	0.101	0.50	0.132	0.31	0.37	0.097	38.8	10.20	0.984
I Feed	--	--	1.149	1.000	--	0.425	1.000	--	1.747	--	10.37	1.000
Bed	14.6	2.62	0.382	0.390	0.72	0.105	0.392	2.44	0.356	46.97	6.86	0.833
Cycl	8.5	5.57	0.473	0.483	1.38	0.117	0.436	10.38	0.882	15.48	1.32	0.160
Bagh.	1.32	9.52	0.125	0.127	3.48	0.046	0.172	16.86	0.222	4.64	0.061	0.007
I Prod.	--	--	0.980	1.000	--	0.268	1.000	--	1.460	--	8.24	1.00

Table 12.3. Elemental analyses and element balances, Run 830226. Operating conditions: Minto coal, 6.6 mm; Havelock limestone, 1.72 mm; excess air = 17.2 %; Ca/S = 4.47; τ = 0.61; v = 293 m/s; T_b = 927 °C.

	$\dot{m}_g$ kg/hr	Element: Si				Element: Al				Element: Fe				Element: Ca			
		w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %				
Coal	39.4	2.65	1.044	0.851	0.75	0.296	0.59	4.23	1.666	0.916	0.43	0.169	0.010				
Stone	41.4	0.44	0.182	0.149	0.50	0.207	0.41	0.37	0.153	0.084	38.8	16.06	0.990				
i Feed	--	--	1.226	1.000	--	0.503	1.00	--	1.819	1.000	--	16.23	1.000				
Bed	19.2	2.01	0.386	0.346	0.35	0.067	0.295	1.71	0.328	0.201	31.4	6.03	0.680				
Cycl.	13.0	4.70	0.611	0.547	0.95	0.124	0.546	8.35	1.086	0.667	20.5	2.66	0.300				
Bagh.	1.6	7.51	0.120	0.107	2.25	0.036	0.159	13.44	0.215	0.132	10.8	0.173	0.019				
i Prod.	--	--	1.117	1.000	--	0.227	1.000	--	1.629	1.000	--	8.86	1.000				

Table 12.4. Elemental analyses and element balances, Run 830316. Operating conditions: Devco coal, 6.6 mm; Havelock limestone, 1.72 mm; excess air, 43.6 %; Ca/S = 2.27; $r = 0.76$; $v = 2.67$ m/s; $T_b = 860$ °C.

s	$\dot{m}_g$ kg/hr	Element: Si			Element: Al			Element: Fe			Element: V		
		w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i ppm	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %
Coal	35.2	3.88	1.366	0.959	2.52	0.887	0.931	3.32	1.168	0.960	7	0.00025	0.63
Stone	13.2	0.44	0.058	0.041	0.50	0.066	0.069	0.37	0.049	0.040	11	0.00015	0.37
i Feed	--	--	1.424	1.000	--	0.953	1.000	--	1.217	1.000	--	0.00040	1.00
Bed	10.6	5.63	0.597	0.524	2.77	0.293	0.509	1.58	0.167	0.171	21	0.00022	0.30
Cycl.	5.0	8.60	0.430	0.378	4.30	0.215	0.373	13.12	0.656	0.670	61	0.00031	0.42
Bagh.	1.14	9.74	0.111	0.098	5.98	0.068	0.118	13.69	0.156	0.159	188	0.00021	0.28
i Prod.	--	--	1.138	1.000	--	0.576	1.000	--	0.979	1.000	--	0.00074	1.0000
		Element: Ca			Element: Mg			Element: Na			Element: K		
		w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i %	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %	w_i ppm	$\dot{m}_{i,s}$ kg/hr	$\phi_{i,s}$ %
Coal	35.2	0.14	0.049	0.010	0.006	0.0021	0.07	0.02	0.007	0.47	0.01	0.004	0.13
Stone	13.2	38.8	5.12	0.990	0.22	0.029	0.93	0.06	0.008	0.53	0.21	0.028	0.87
i Feed	--	--	5.17	1.000	--	0.0311	1.00	--	0.015	1.00	--	0.032	1.00
Bed	10.6	20.36	2.16	0.976	0.21	0.022	0.737	0.02	0.002	0.16	0.02	0.002	0.12
Cycl.	5.0	0.95	0.048	0.022	0.12	0.0060	0.201	0.08	0.004	0.32	0.07	0.0035	0.22
Bagh.	1.14	0.47	0.0054	0.0024	0.16	0.0018	0.060	0.57	0.065	0.52	0.94	0.0107	0.66
i Prod.	--	--	2.21	1.000	--	0.0298	1.000	--	0.0125	1.00	--	0.0162	1.00

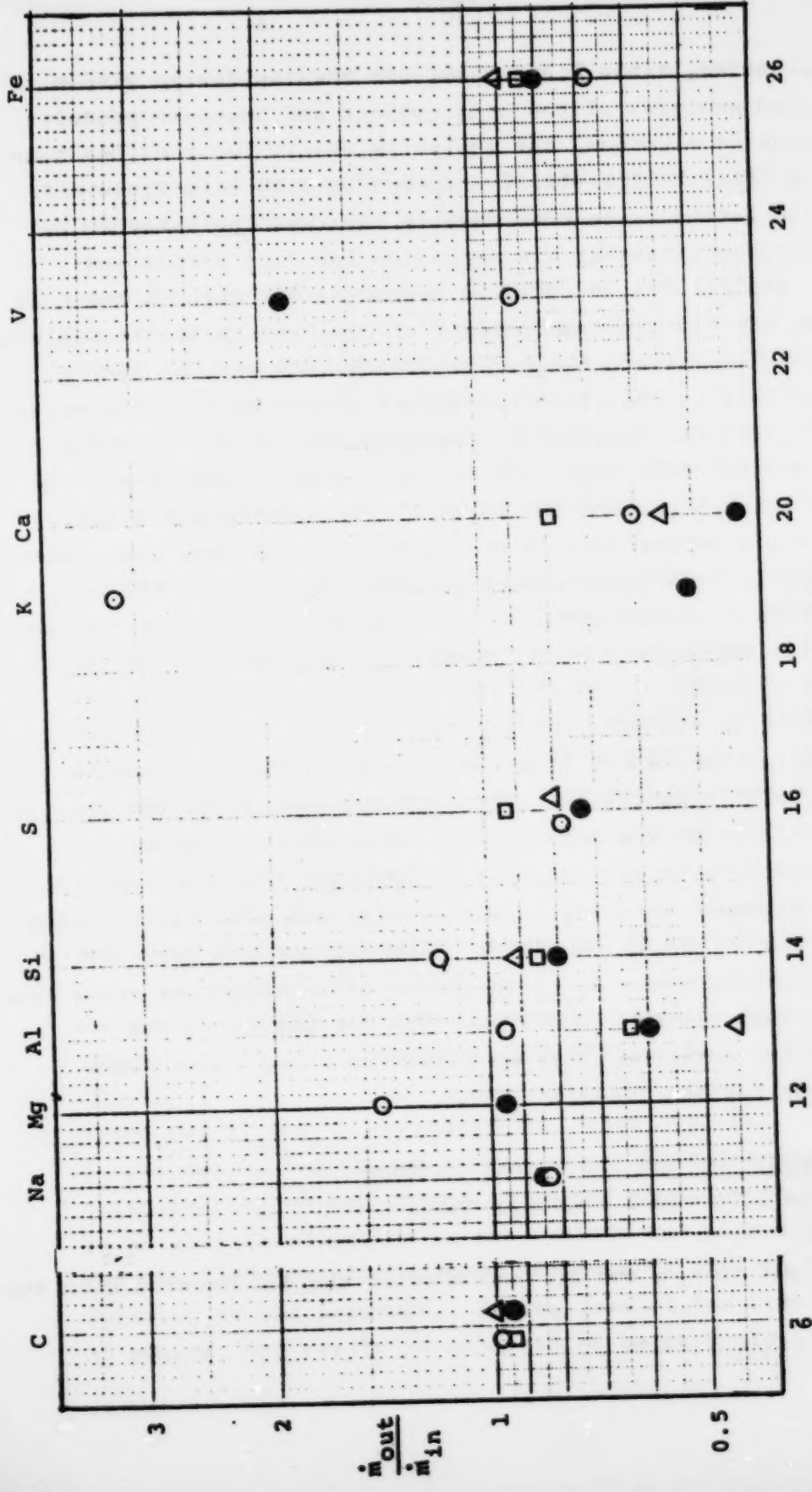


Fig. 12.1. Element balances for C, Na, Mg, Al, Si, S, K, Ca, V, and Fe for four selected runs: the flux ratio of an element in the product streams to that element in the feed streams, vs. atomic number (chosen for convenience, not significance). Runs: $\circ$, 830201; $\square$, 830225; $\triangle$, 830226; $\bullet$, 830316. Coals: $\bullet$, Devco; others, Minto.

the same quarry, Table 4.8. The fault thus evidently lies with the calcium analyses of the bed, cyclone and baghouse products. In the case of aluminum, our analyst's result for the limestone is 0.50 %, Tables 12.1--12.3, compared to 0.08 % in Table 4.8. Use of the latter value considerably improves the balances but still leaves substantial discrepancies for Runs 830226 and 830316. The difference could be accounted for through the coal analysis, but the aluminum content of coal ash is too variable to allow close comparison between results on our samples and published data on coal from the same fields. We can only observe that our analyst's results for the aluminum content of Devco coal are average for that coal, while those on Minto are on the low side but not impossible. However, an explanation which would require the aluminum content of the Minto to be even lower seems unlikely. It thus appears most probable that, as in the case of calcium, the aluminum contents found by our analyst for the bed, cyclone and baghouse products tended to fall well on the low side.

Some of the results reported by our analyst are thus evidently significantly in error, and the quality of the analyses tended to be below the expected and attainable standard. Nevertheless, there are apparent in the data certain trends which are much stronger than any likely errors, and observations upon these should therefore be valid. Also, it appears that the element flux fractions $\phi_{i,st}$ should be less subject to error than the individual fluxes themselves. The discussion in the following two sections takes up inquiries along these lines without any further justification.

12.4 Composition and flux distributions

Tables 12.1--12.3 indicate that:

- o almost all the Fe and Si, and most of the Al, entered with the coal

- o nearly all the Ca and Mg, and most of the K, entered with the limestone
- o the supply of Na and V was relatively evenly divided between coal and limestone
- o among the products, the bed product was the richest in, and was the major carrier of, Ca and Mg. The baghouse product was by far the leanest in these elements, and the smallest carrier. However, it is clear that as total recycle of cyclone dust is approached closely enough ($r \rightarrow 1$), the cyclone product must diminish as a carrier relative to the baghouse product, and disappear altogether at $r = 1$.
- o the bed product was lean in Fe and the baghouse product was rich. As a result, the baghouse product, even when it was a relatively small stream, was a substantial carrier of Fe. However, except at $r \rightarrow 1$, the cyclone product was the major carrier. The richness of the baghouse product in Fe was visually apparent; the material was of a dark greyish-red appearance, like hematite powder intimately mixed with a leaden-grey powder. The cyclone product, by contrast, was simply dark leaden grey.
- o the behaviour of Al was quite similar to that of Fe
- o except in the run at highest recycle (830201, at $r = 0.919$), the behaviour of Si was very like that of Al
- o since Fe, Al and Si represent the major coal-ash constituents, and Ca and Mg represent the major limestone constituents, we see in the above the tendency of coal ash to be elutriated (going to the cyclone and baghouse products), and the tendency of limestone residues to stay in the bed (leaving as bed product). In the case of the limestone, of course, the result is expected to be strongly affected

by the initial content of elutriable fines. In the case of coal, sensitivity to initial particle size distribution must be considerably less since a substantial portion of the ash consists by nature of very fine particles, smaller than those produced by crushing of the coal and virtually unaffected by that crushing.

- o the baghouse product, interestingly, was the major out-carrier of Na and K even though coal was the major in-carrier.
- o vanadium, which entered almost equally distributed between coal and limestone, was most concentrated in the baghouse product, and very strongly so in the run with Devco coal. The transport out was however relatively evenly divided between the bed, cyclone and baghouse products, because of the relative volumes of the streams.

12.5 Further comparisons of element flux distributions and effects of recycle

The element flux distributions, at least on the evidence of the few runs here considered, seem to depend strongly and quite regularly on the recycle fraction r , Figs. 12.2--12.4. If this is indeed so, then they depend only weakly, relatively, on the other factors that varied between the subject runs. Lacking other evidence or any contrary indications, we take this interpretation as a working hypothesis in the following discussion.

The flux fraction taken by the cyclone product must drop to zero as $r \rightarrow 1$. That is all that we know a priori about the limit behaviour in Figs. 12.2--12.4; the rest has been filled in by guesswork.

In Fig. 12.2 we see qualitative similarities between the behaviours of Fe and Al, but also some divergence. The flux fractions carried by the baghouse product are very nearly the

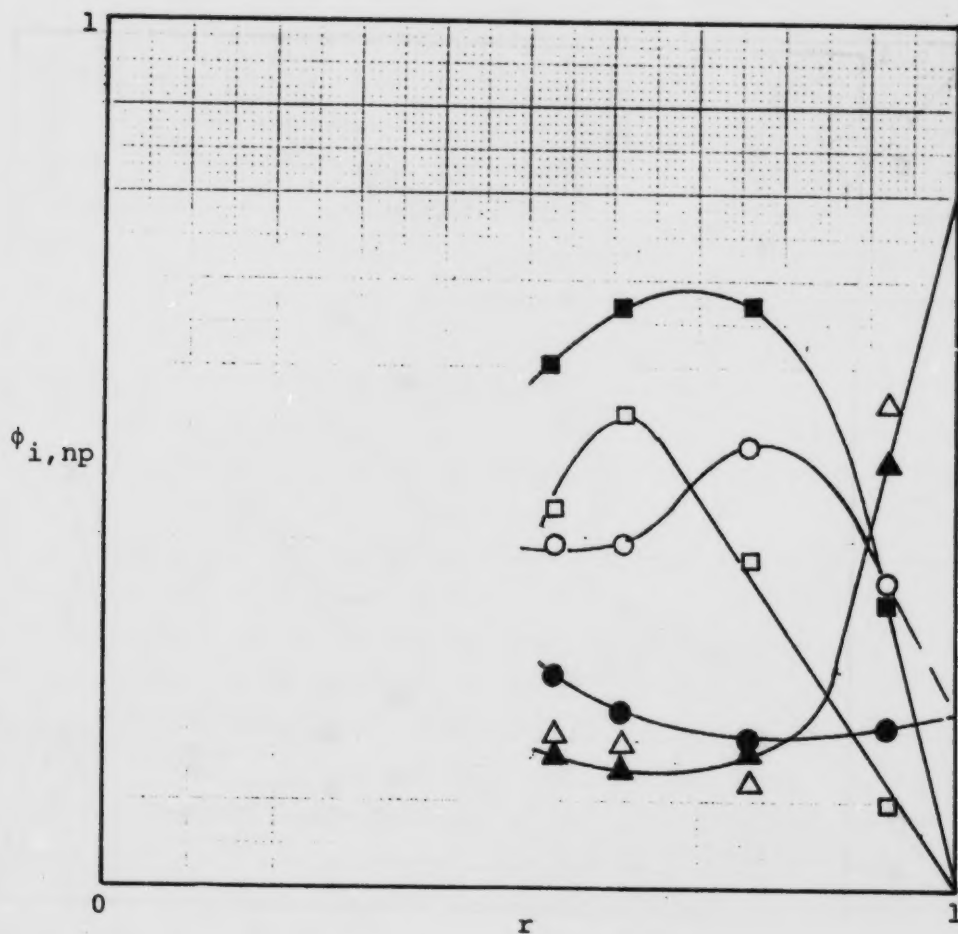


Fig. 12.2. Partitioning of Al and Fe element fluxes between the product streams at various recycle fractions. Element: $\circ \square \triangle$, Al; $\bullet \blacksquare \blacktriangle$, Fe. Product: $\circ \bullet$, bed; $\square \blacksquare$, cyclone; $\triangle \blacktriangle$, baghouse.

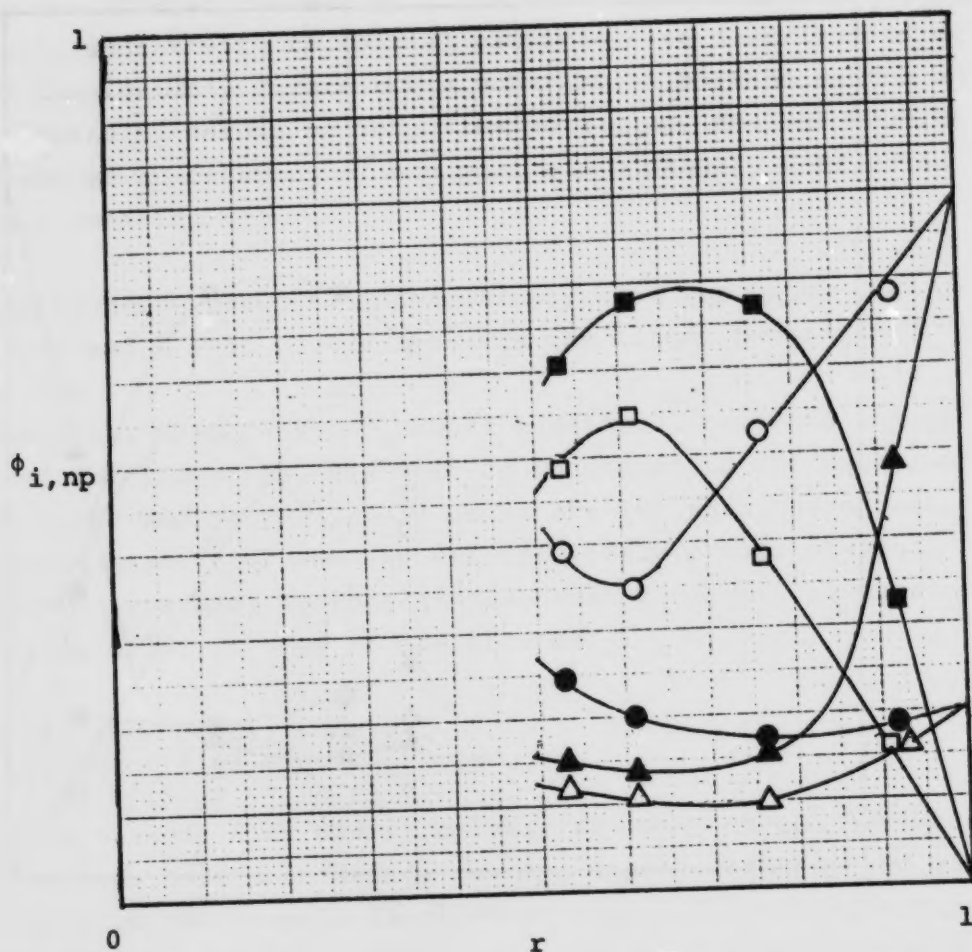


Fig. 12.3. Partitioning of Si and Fe element fluxes between the product streams at various recycle fractions. Element: $\circ \square \triangle$, Si; $\bullet \blacksquare \blacktriangle$, Fe. Product: see key in Fig. 12.2.

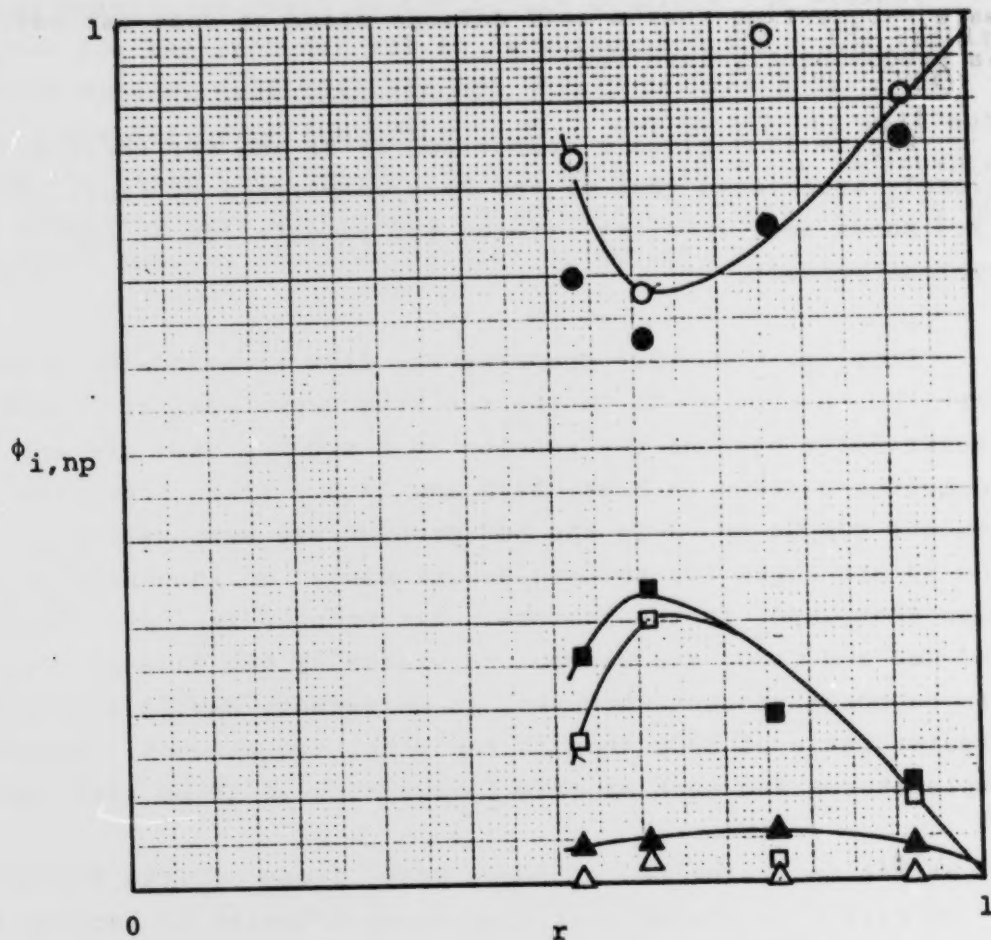


Fig. 12.4. Partitioning of Ca and S element fluxes between the product streams at various recycle fractions. Element: $\circ \square \triangle$, Ca; $\bullet \blacksquare \blacktriangle$, S. Product: see key in Fig. 12.2.

same, suggesting that a well defined class of coal-ash particles is predominantly involved.

Figure 12.3 gives a comparison of Si and Fe; these are less closely associated than Al and Fe. Comparison of Figs. 12.2 and 12.3 shows near identity between the results for Al and Si in the cyclone product.

Finally, Fig. 12.4 compares the flux fractions of S and Ca. Here the results on Ca in the run with Devco coal are anomalous, rather oddly because the results on S are not, nor are any such anomalies visible in Figs. 12.2 and 12.3. Apart from this, the results for Ca and S in the bed product are very nearly the same, and so are those for the cyclone product. We interpret this to mean that CaSO_4 is the dominant carrier of S in these streams, and CaO and CaSO_4 are present in virtually the same proportions. The baghouse product, however, is different, having a very much greater ratio of S/Ca than in the other two streams. Possible explanations for such an effect, assuming it to be real, are two:

- o attrition in the bed produces CaSO_4 fines off the surface of partly sulphated particles whose interior is largely still CaO. Elutriable fines entering in the limestone feed are quickly and fully sulphated. Thus the finest fraction of limestone residues--that in the baghouse product--tends to be the richest in CaSO_4 .
- o char in the baghouse material is, possibly, richer in S than char in the cyclone product. However, since such char constituted only 5--17 % of the baghouse product in the subject runs, this cannot have been the major sulphur-enriching factor.

12.6 Closure

We emphasize that the results just presented, based on only four runs, must be treated with appropriate caution. If time and

funding permit, we shall in the future complete the element analyses for the runs not yet so favoured, and check the results on those already done that are not wholly satisfactory. Very likely, however, such further work will add detail but will not change the basic conclusions reached here. It seems improbable that the reasonably consistent and coherent picture given by the considered runs is uncharacteristic or accidental.

13. CHARACTERISTICS OF THE SLOW
FLUCTUATIONS IN BED TEMPERATURE,
GAS COMPOSITION, AND SOLIDS FEED RATES

13.1 Introduction

Attention was given in Sect. 5.5 to the slow fluctuations of variables, with time constants typically in the range 10 min.-- 3 hr, which exist in all states of operation, including the steady state. These fluctuations will now be examined statistically, to characterize them as to intensity (i.e., characteristic amplitude, measured by the root-mean-square deviation or "standard error"), and to search for any correlations that may exist between them. It will be seen that in some cases the correlations are highly suggestive as to the effects of variations in certain process variables. The results also have interesting implications regarding process control.

13.2 Intensity of fluctuations in
bed temperature and combustor-exit
gas composition

We have selected for examination eight runs on which full and long records of bed temperature and combustor-exit gas composition were obtained. In some of the other runs the record of exit gas composition, for example, was broken for a considerable periods of time while traverses of the freeboard were carried out. Table 13.1 gives, for the chosen eight runs, the results on the coal feedrate, the bed temperature, and the content in the gases at combustor exit of O_2 , NO_x and SO_2 . It was concluded in Sect. 5.4 that bed temperature and exit gas composition effectively reach steady-state levels (more accurately, the statistically stationary state) relatively early. The results in Table 13.1 are based on the records at COT time $t > \tau_p = m_b / \dot{m}_p$.

Table 13.1. Intensity of slow fluctuations in bed temperature and in gas composition at combustion exit, $t > t_{ps}$.

Run	D _p , mm		v, m/s	ε _a , %	Ca/S,	m _f , kg/hr	T _b , °C	X _{O₂} , %	X _{NO_x} , ppm	X _{SO₂} , ppm	τ
	coal	stone									
830215	6.6	0.78	2.77	14.7	2.24	41.4 ± 1.4 (± 3.5 %)	851 ± 21 (± 2.5 %)	3.58 ± 0.69 (± 3.7 %)	298 ± 33 (± 2.0 %)	1210 ± 150 (± 2.6 %)	0.40
830216	6.6	0.78	2.76	14.9	2.86	41.6 ± 1.0 (± 2.5 %)	855 ± 12 (± 1.4 %)	3.60 ± 0.16 (± 0.9 %)	323 ± 9 (± 0.9 %)	750 ± 140 (± 2.2 %)	0.48
830217	6.6	0.78	2.68	17.5	3.08	38.0 ± 0.75 (± 2.0 %)	849 ± 10 (± 1.2 %)	4.64 ± 0.64 (± 3.6 %)	363 ± 10 (± 0.6 %)	710 ± 70 (± 1.3 %)	1
830218	19.3	1.98	2.77	24.0	2.70	37.6 ± 1.33 (± 3.5 %)	860 ± 19 (± 2.3 %)	6.16 ± 0.60 (± 3.6 %)	353 ± 18 (± 1.1 %)	2900 ± 590 (± 11.4 %)	1
830224	19.3	1.72	2.68	21.8	3.56	38.0 ± 1.4 (± 3.7 %)	859 ± 11 (± 1.3 %)	4.33 ± 0.75 (± 4.5 %)	335 ± 23 (± 1.5 %)	160 ± 220 (± 4.1 %)	0.43
830225	6.6	1.72	2.98	19.3	3.10	39.0 ± 0.9 (± 2.4 %)	943 ± 13 (± 1.4 %)	4.10 ± 1.33 (± 7.7 %)	297 ± 39 (± 2.7 %)	2170 ± 690 (± 12.7 %)	0.47
830226	6.6	1.72	2.93	17.2	4.47	39.4 ± 1.1 (± 2.8 %)	927 ± 11 (± 1.2 %)	2.86 ± 0.87 (± 5.0 %)	295 ± 48 (± 2.7 %)	640 ± 320 (± 5.8 %)	0.39
830322	19.3	0.49	1.82	1.2	2.14	32.3 ± 1.9 (± 6.0 %)	847 ± 17 (± 2.0 %)	4.10 ± 0.52 (± 2.6 %)	290 ± 21 (± 1.2 %)	1560 ± 220 (± 3.3 %)	1

t) rather than the feedrate itself. The results indicate that the fluctuation intensity (the r.m.s. amplitude, or the "standard error") was as low as 2 %, as high as 6 %, and usually around 3 %.

The fluctuation intensity of bed temperature was 10--20 °C, and typically around 15 °C. Measured relative to the temperature rise $T_b - T_o$, this represents a fluctuation level of 1.2--2.5 %, and typically 1.5 %, where T_o is characteristic of the feed-stream temperatures and is here taken to be the inlet air temperature. These fluctuation levels are 1/3--2/3 those estimated for the coal feedrate. If variability in the coal feedrate is the controlling factor in the slow fluctuations of temperature and gas composition, then the bed temperature fluctuations are indeed expected to be damped relative to those in the coal feedrate, because of the thermal inertia of the bed and the slowness of combustion of large char particles. Furthermore, the damping observed is of a magnitude commensurate with this explanation.

For comparison with each other and with the fluctuations in the fuel flux m_f and the bed temperature elevation $T_b - T_o$, the fluctuations of O_2 , NO_x and SO_2 in the gases at combustor exit have been normalized in Table 13.1 as follows:

- o for O_2 : (fluctuation in O_2 efflux)/(O_2 influx theoretically required to combust the coal)
- o for NO_x : (fluctuation in molar efflux of NO_x)/(molar influx of fuel N)
- o for SO_2 : (fluctuation in molar efflux of SO_2)/(molar influx of fuel S)

The table indicates that in these terms the fluctuations in O_2 content of the gases at combustor exit were usually similar to

those in the coal flow, but sometimes two or three times as large. The NO_x fluctuations were generally smaller than those in the coal flow and usually smaller than those in temperature. The SO_2 fluctuations were relatively quite high in a few cases, but in most they were within a factor of two of those in the coal flow.

It was noted in Sect. 5 that the fluctuations in air flow were negligible relative to those in the coal. The third feed stream, the limestone, however, underwent fluctuations similar to those in the coal. There is no evidence of correlation between the fluctuations of the coal and limestone flows, nor is any expected under our operating conditions. The effect of the fluctuation in limestone flow, added to the uncorrelated fluctuation in the coal flow, should therefore be to increase the fluctuations in bed temperature and gas composition, likely in rough accordance with the rule for addition of random errors; e.g., in the case of temperature fluctuations,

$$s_T^2 = a^2 s_f^2 + b^2 s_s^2 \quad (13.2.1)$$

where s_T is the "standard error" (r.m.s. amplitude) of the temperature fluctuations, s_f and s_s are the "standard errors" of the fluctuations in the coal and limestone flows, and a^2 and b^2 are weighting factors. The ratio a/b in this case can be roughly estimated from knowledge of the reaction thermodynamics. The specific enthalpy of combustion of Minto coal is about - 30 MJ/kg. The specific enthalpy of calcination of Havelock limestone is about the same as for CaCO_3 , 5.8 MJ/kg. The mass feedrate of the limestone was never greater than that of the coal. It may thus be argued that $a/b > 30/5.8 = 5$, or $a^2/b^2 > 25$. This suggests, according to (13.2.1), that the effect of the limestone on the temperature fluctuations was negligible. We have here ignored the complication of sulphation of the calcined limestone (exothermic) which would further increase the ratio a/b by lowering the limestone weighting b .

The effect of the limestone on the fluctuations in NO_x , if any, is less direct than that on bed temperature, and cannot be gauged so easily. The effect on SO_2 , on the other hand, is rather different and more obvious. The coal is the carrier of sulphur into the system, so it is directly involved. The calcined limestone, on the other hand, is the receptor for SO_2 , so it too is involved. It seems reasonable then to hypothesize that

$$s_{\text{SO}_2}^i = a^i s_f^i + b^i s_s^i \quad (13.2.2)$$

The relation between the effects of coal and limestone is simply chemical (i.e., having to do with the reaction stoichiometry) rather than, as in the case of temperature, thermochemical. It appears that here a/b may be near unity, so it would not be surprising if the limestone sometimes contributed significantly to the fluctuations in SO_2 level. Table 13.1 shows that the normalized SO_2 fluctuations were indeed generally larger than those in temperature, and in some cases very considerably so.

In summary, the evidence of the fluctuation amplitudes alone, Table 13.1, indicates that fluctuations in bed temperature and exit O_2 flux can be reasonably attributed to the variations in coal feedrate alone. This is true as well of the exit flux of NO_x . The fluctuations in SO_2 emission, however, may be significantly augmented by variations in limestone feedrate. The arguments presented, of course, leave out of consideration a variety of possibly important complications and effects which cannot be simply quantified or assessed--e.g., freeboard reactions. Some of these complications and effects are evident in the results as presented in the following section.

13.3 Two-variable correlations between fluctuations

Figure 13.1 shows the relation between bed temperature and exit-gas oxygen content, and Figs. 13.2--13.9 show the relations between the exit-gas contents of $\{\text{CO}, \text{SO}_2, \text{NO}_x, \text{HC}\}$ and the

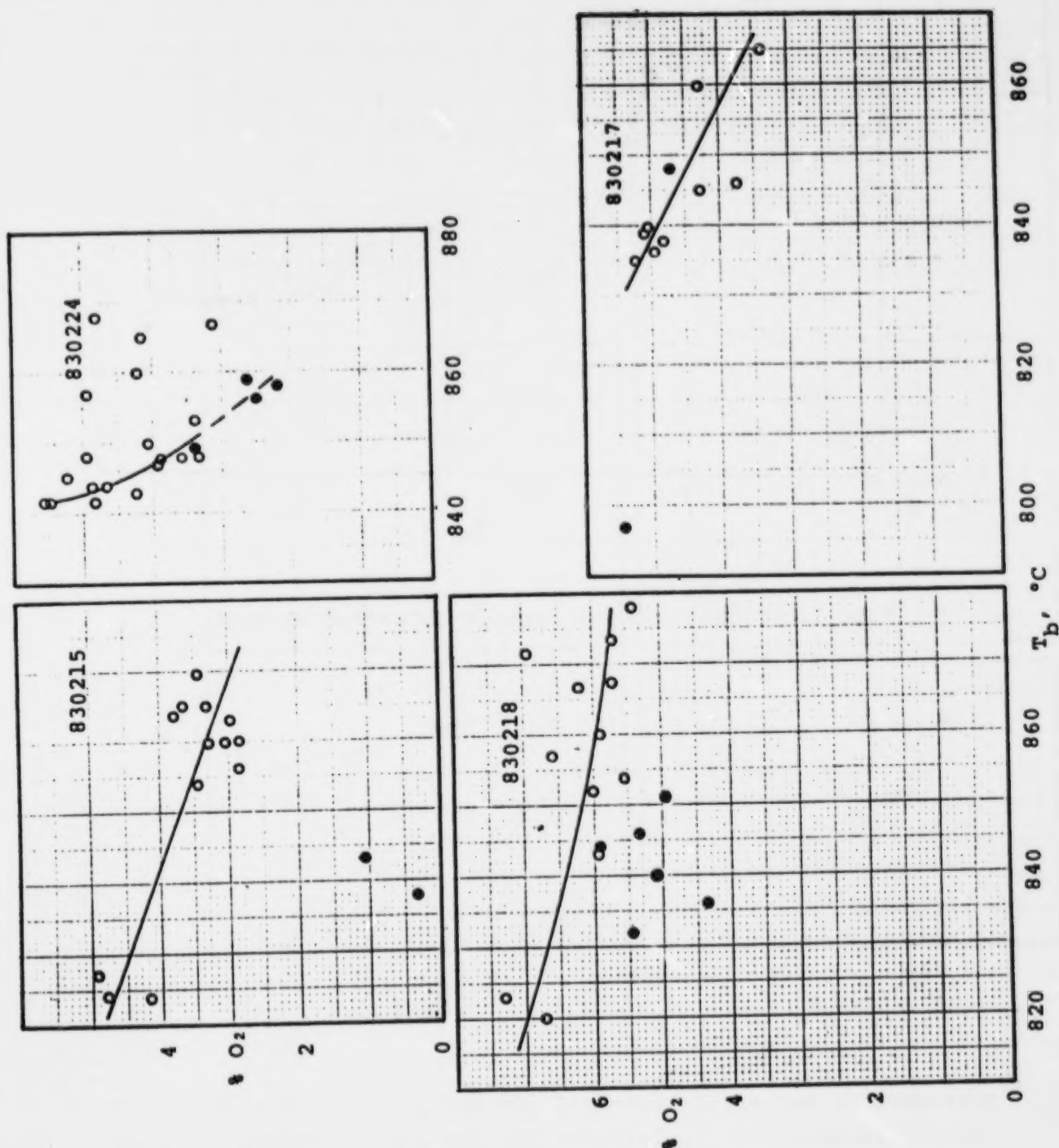


Fig. 13.1. Oxygen content of gas at combustor exit (% by volume, dry basis) vs. bed temperature. Shaded points: $COT < \tau_p$. See Fig. 13.7 for further distinction of points in run 830226.

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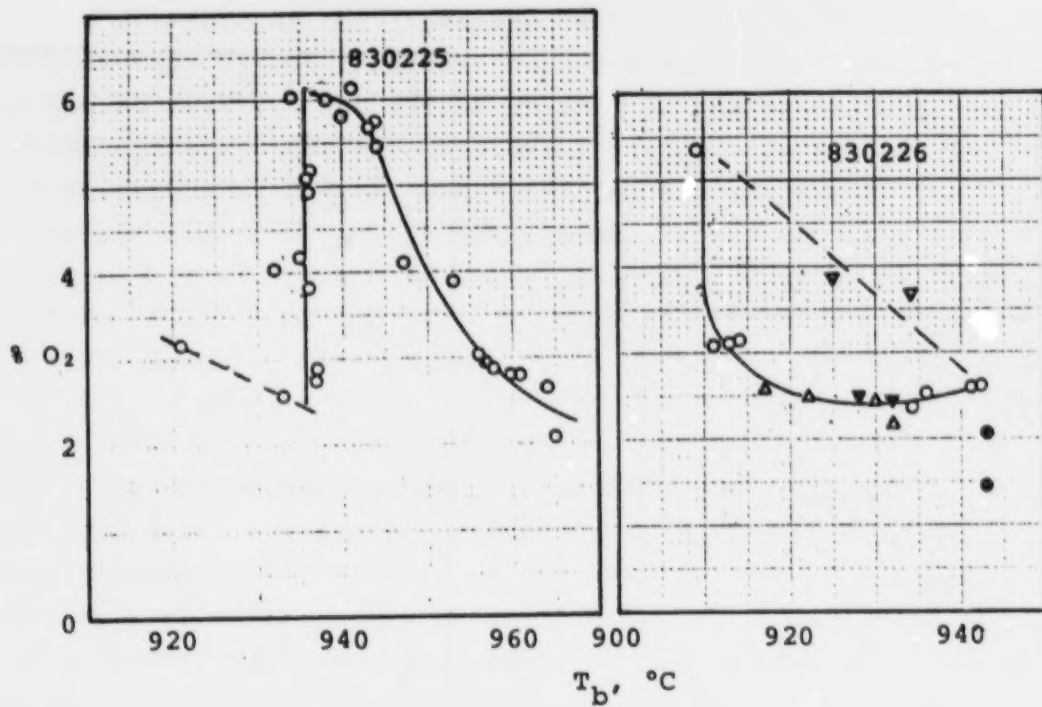


Fig. 13.1, cont'd. See caption on preceding page

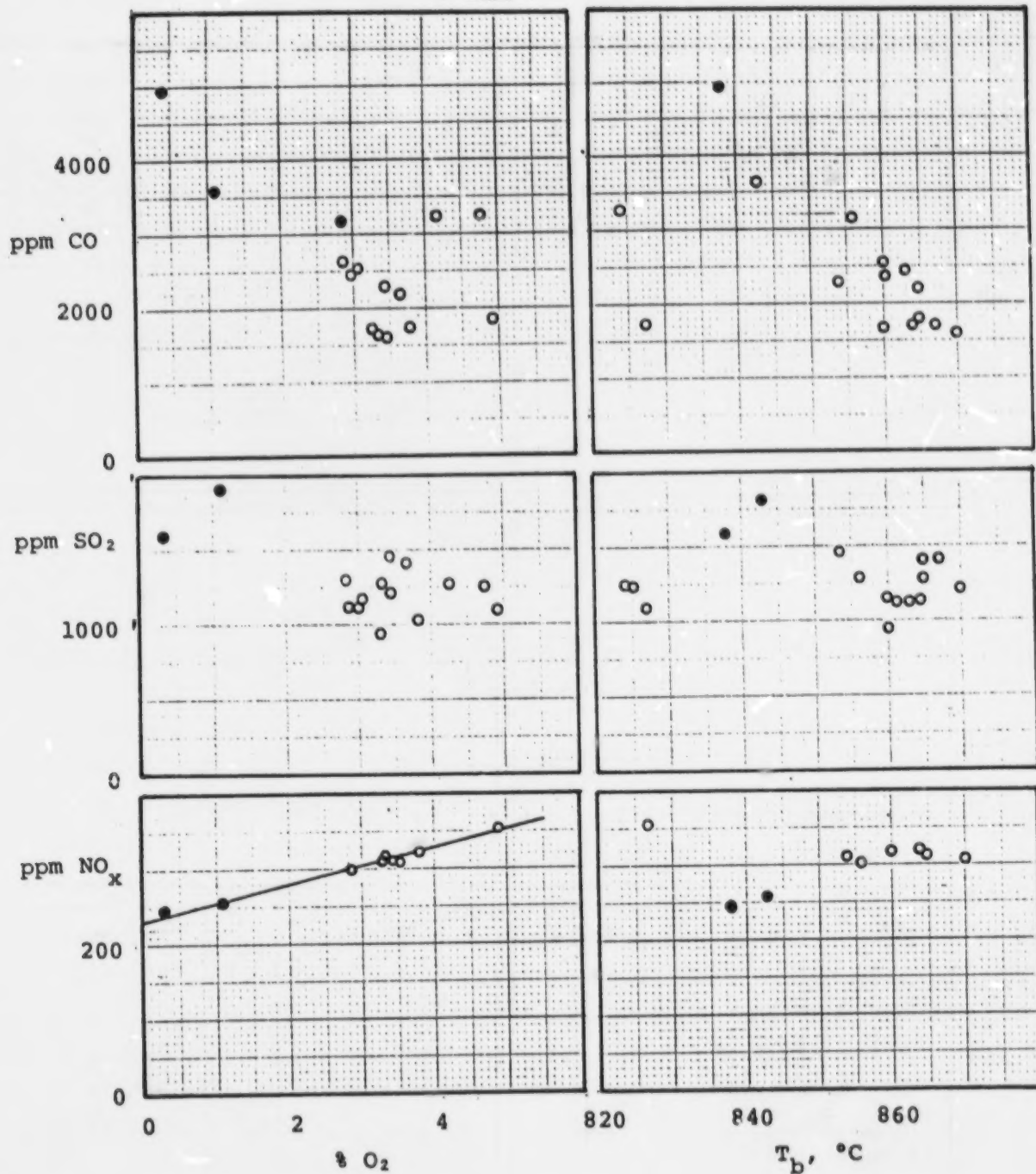


Fig. 13.2. The CO, SO₂ and NO_x contents of the gas at combustor exit vs. oxygen content and bed temperature (gas composition on dry basis), run 830215. Shaded points: COT < τ_p.

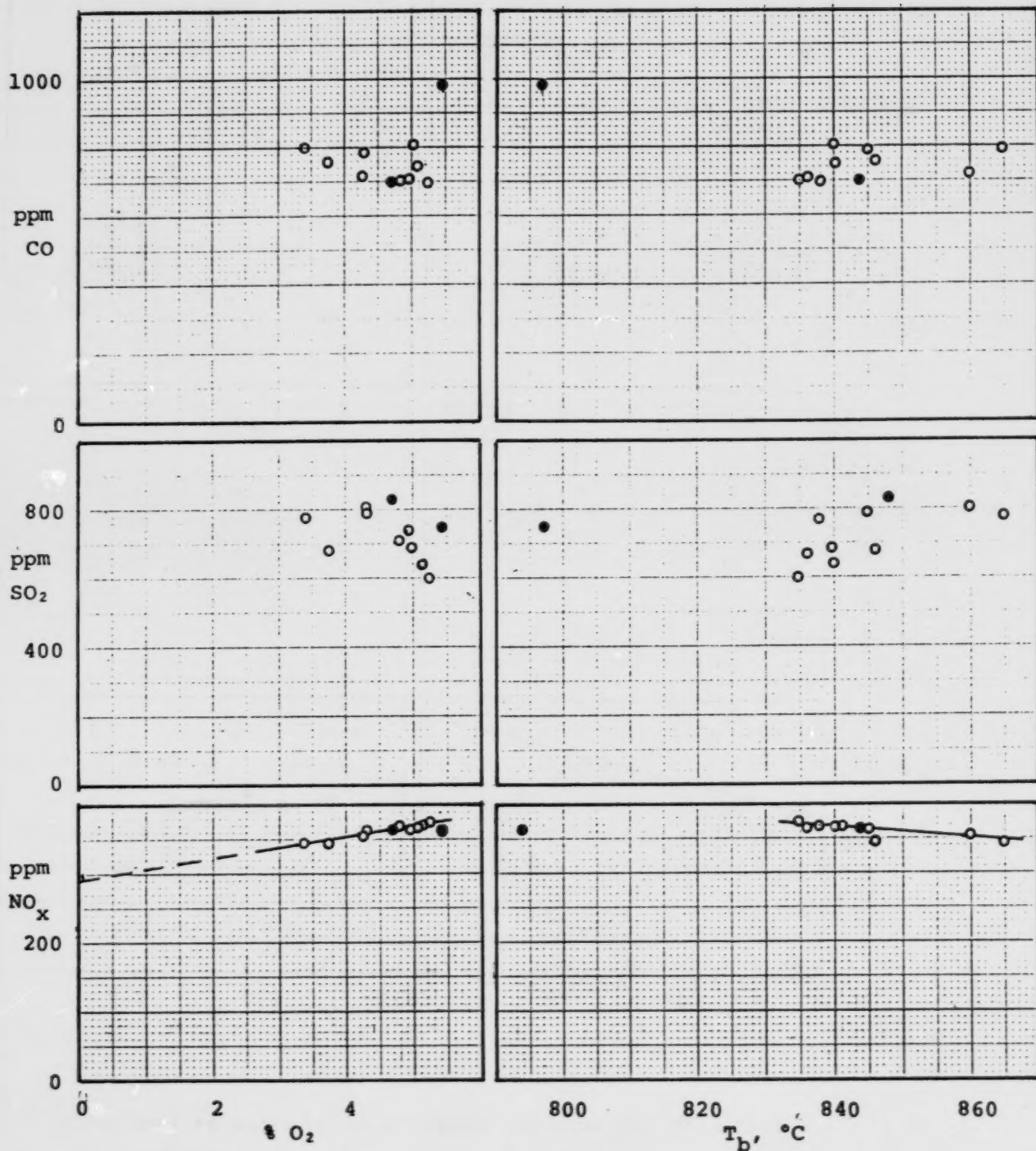


Fig. 13.3. The CO, SO₂ and NO_x contents of the gas at combustor exit vs. oxygen content and bed temperature (gas composition on dry basis), run 830217. Shaded points: COT < τ_p.

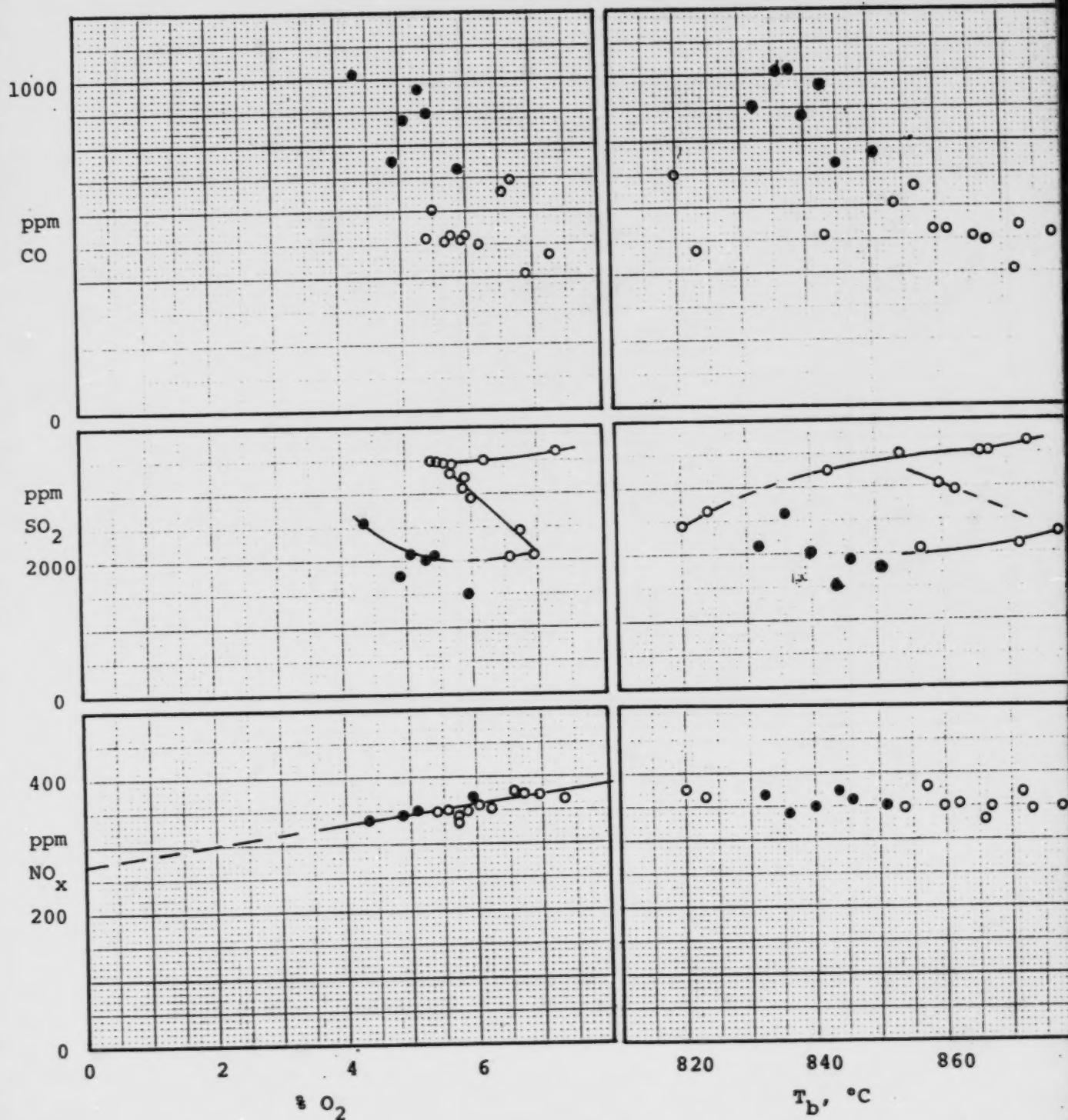
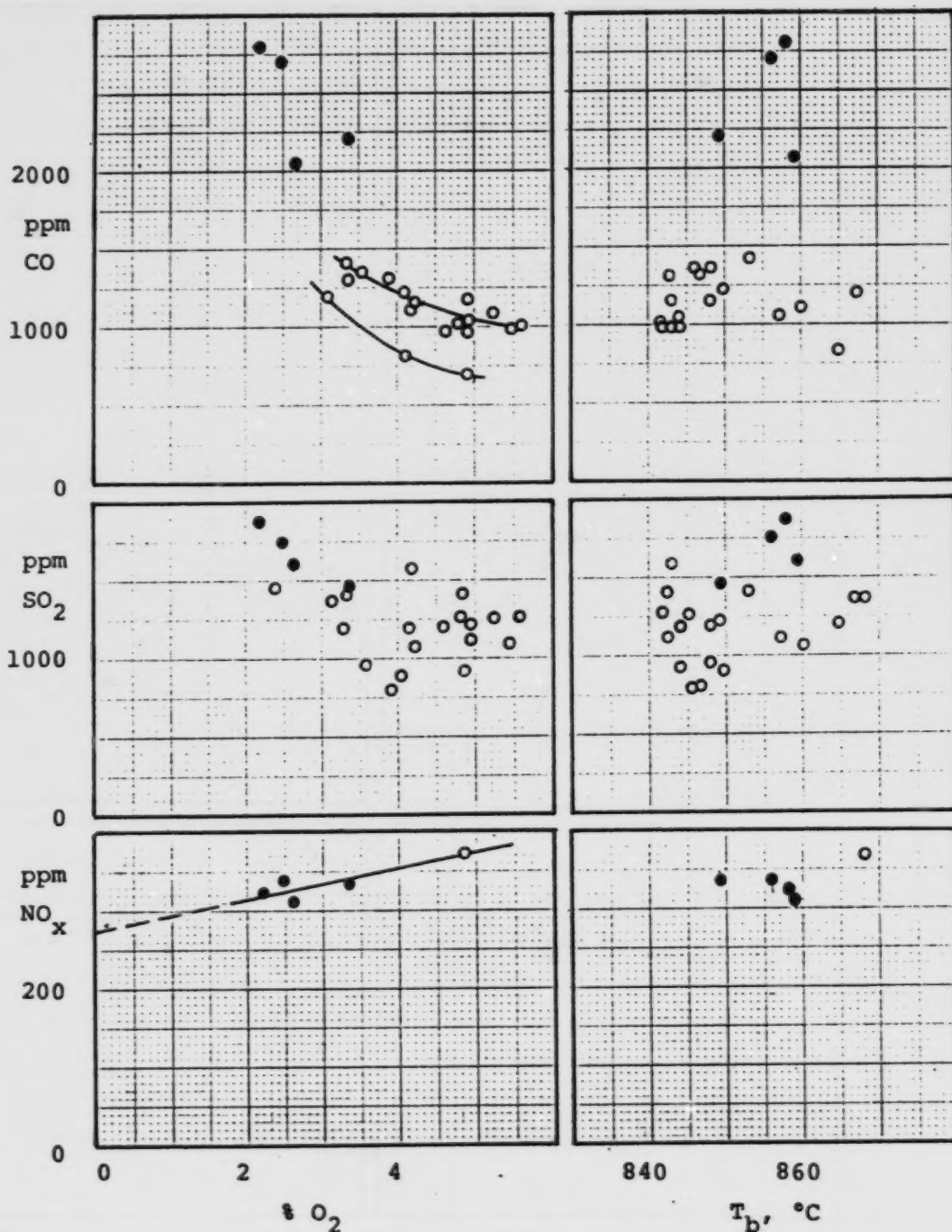


Fig. 13.4. The CO, SO₂ and NO_x contents of the gas at combustor exit vs. oxygen content and bed temperature (gas composition on dry basis), run 830218. Shaded points: COT < τ_p.



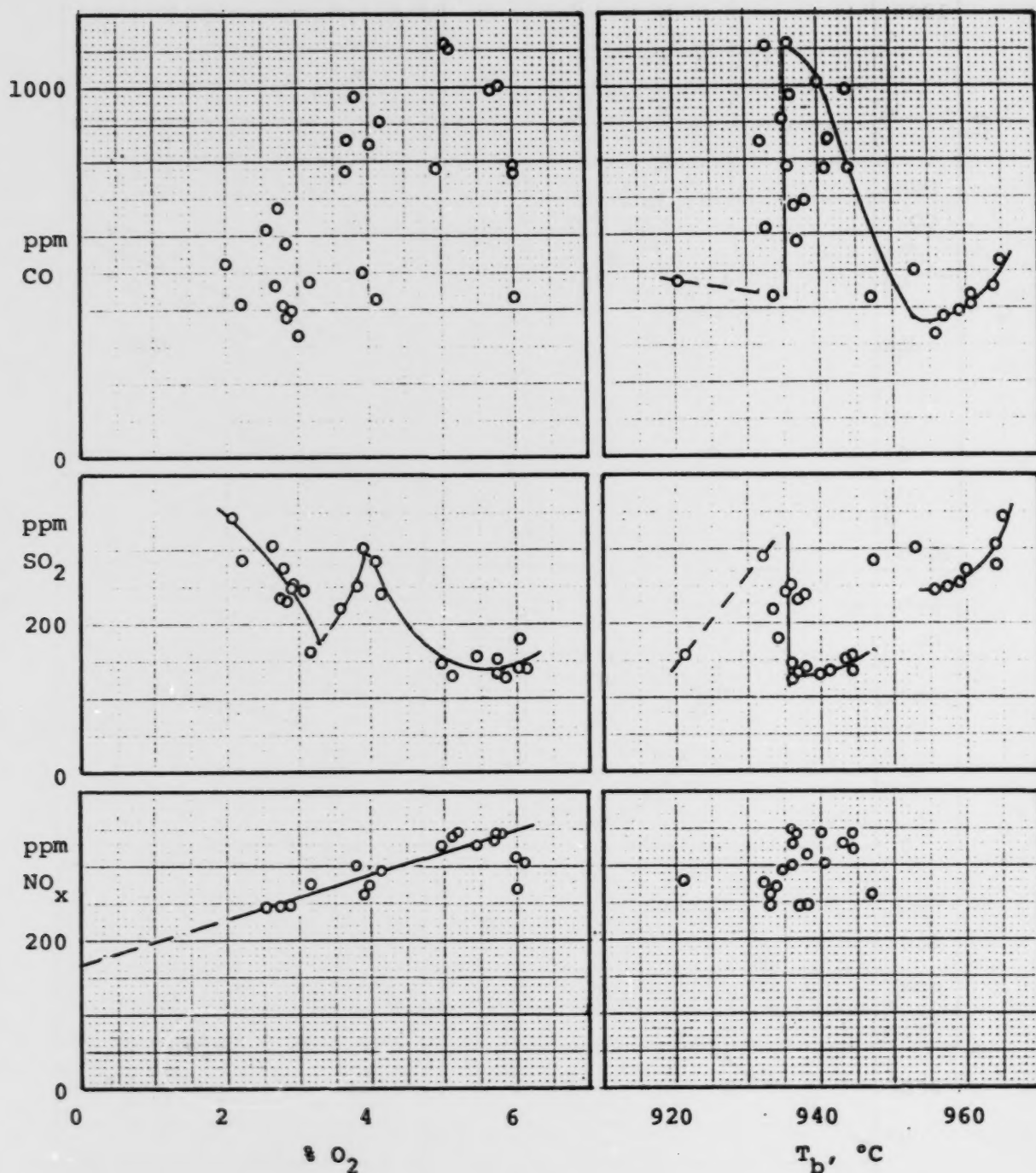


Fig. 13.6. The CO, SO₂ and NO_x contents of the gas at combustor exit vs. oxygen content and bed temperature (gas composition on dry basis), run 830225. Shaded points: COT < τ_p.

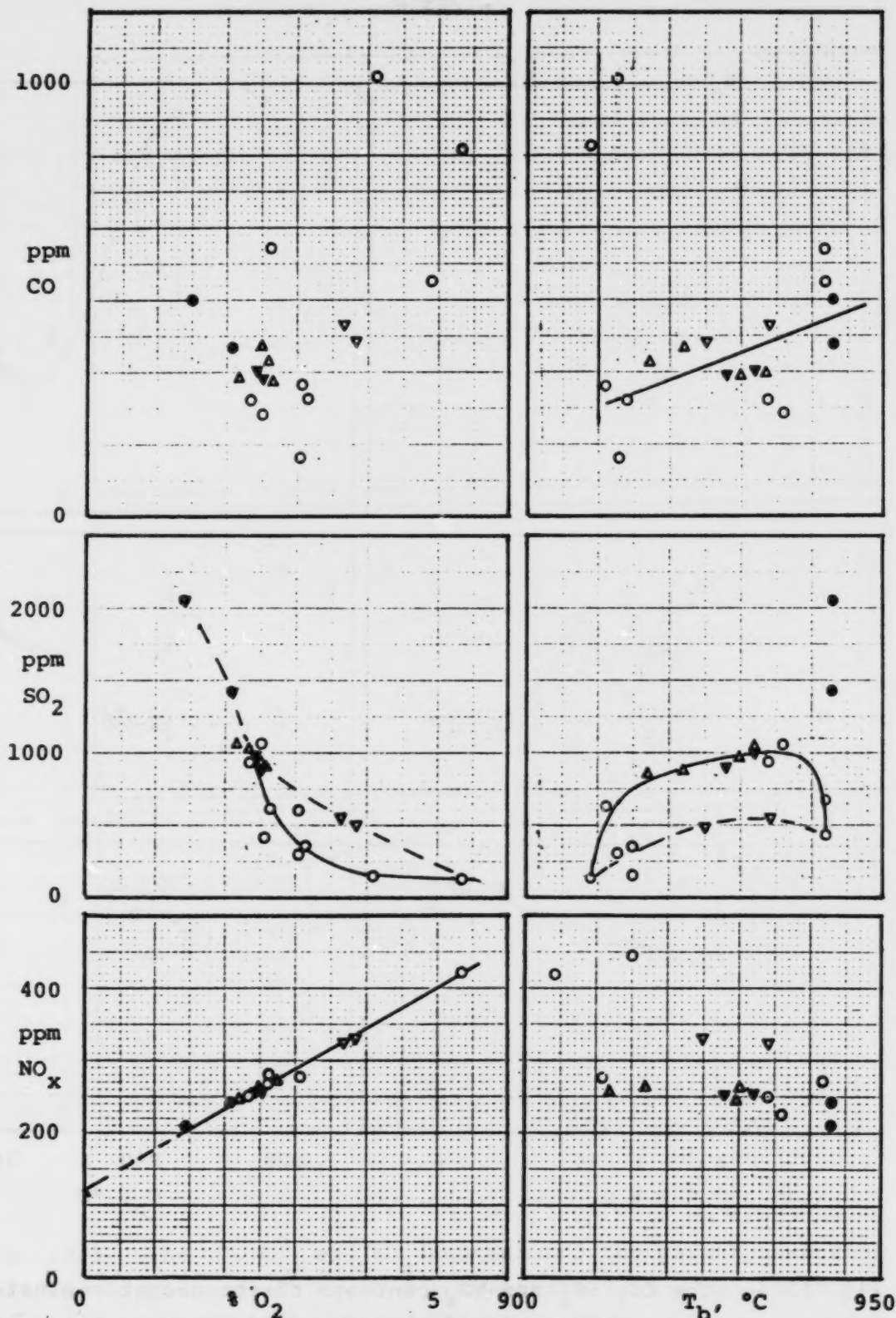


Fig. 13.7. The CO, SO₂ and NO_x contents of the gas at combustor exit vs. oxygen content and bed temperature (gas composition on dry basis), run 830226. Shaded points: COT < τ_p. Further distinctions: Δ▲, T_b rising fairly to very sharply; ∇∇, T_b falling fairly to very sharply; ○●, T_b at extremum or changing

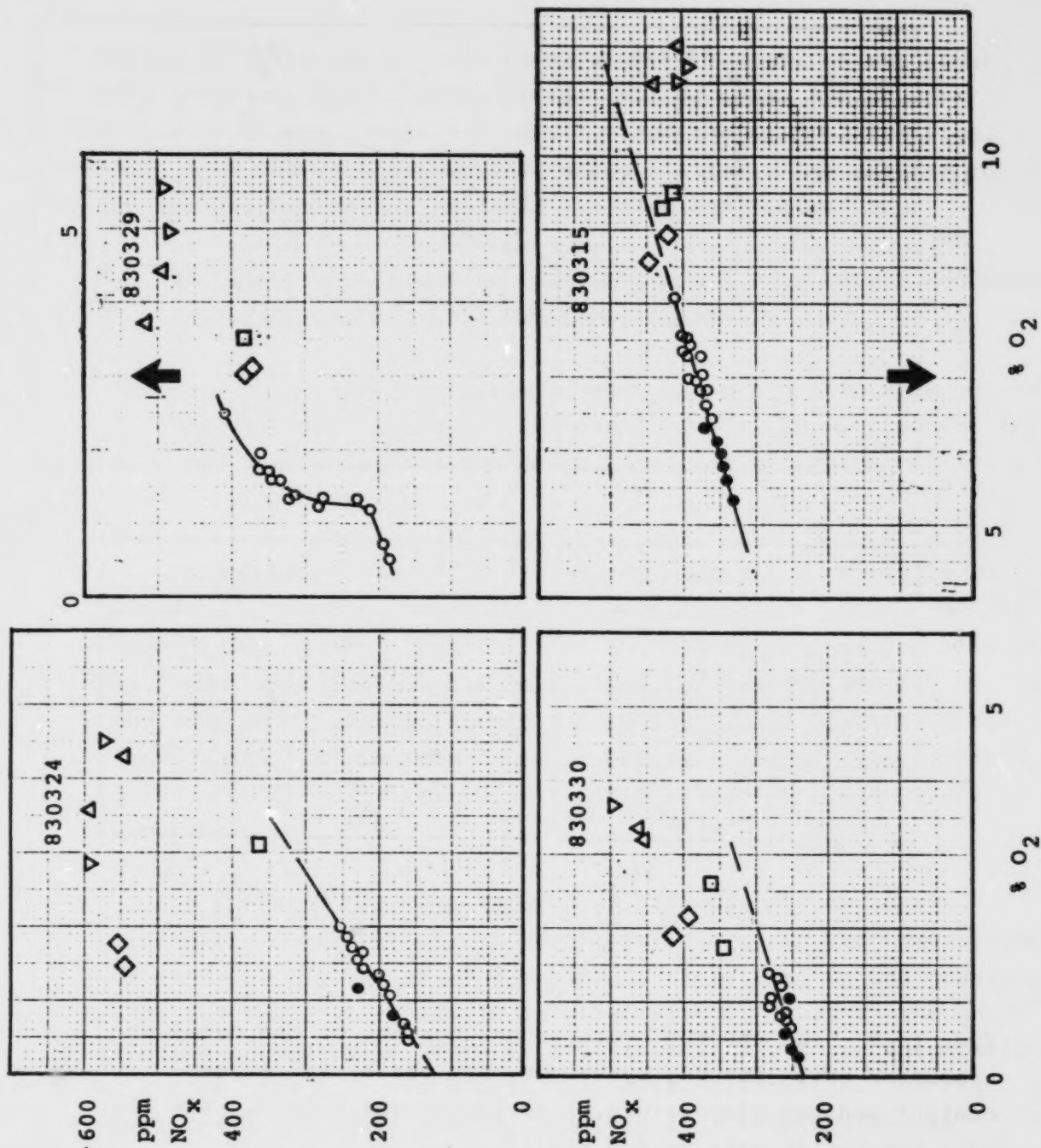


Fig. 13.8. The NO_x content of the gas at various levels in the freeboard vs. the oxygen content. Height above distributor plate, m: □, 3.24; ◇, 2.60; △, 1.41; ▽, 1.11; rest, at combustor exit.

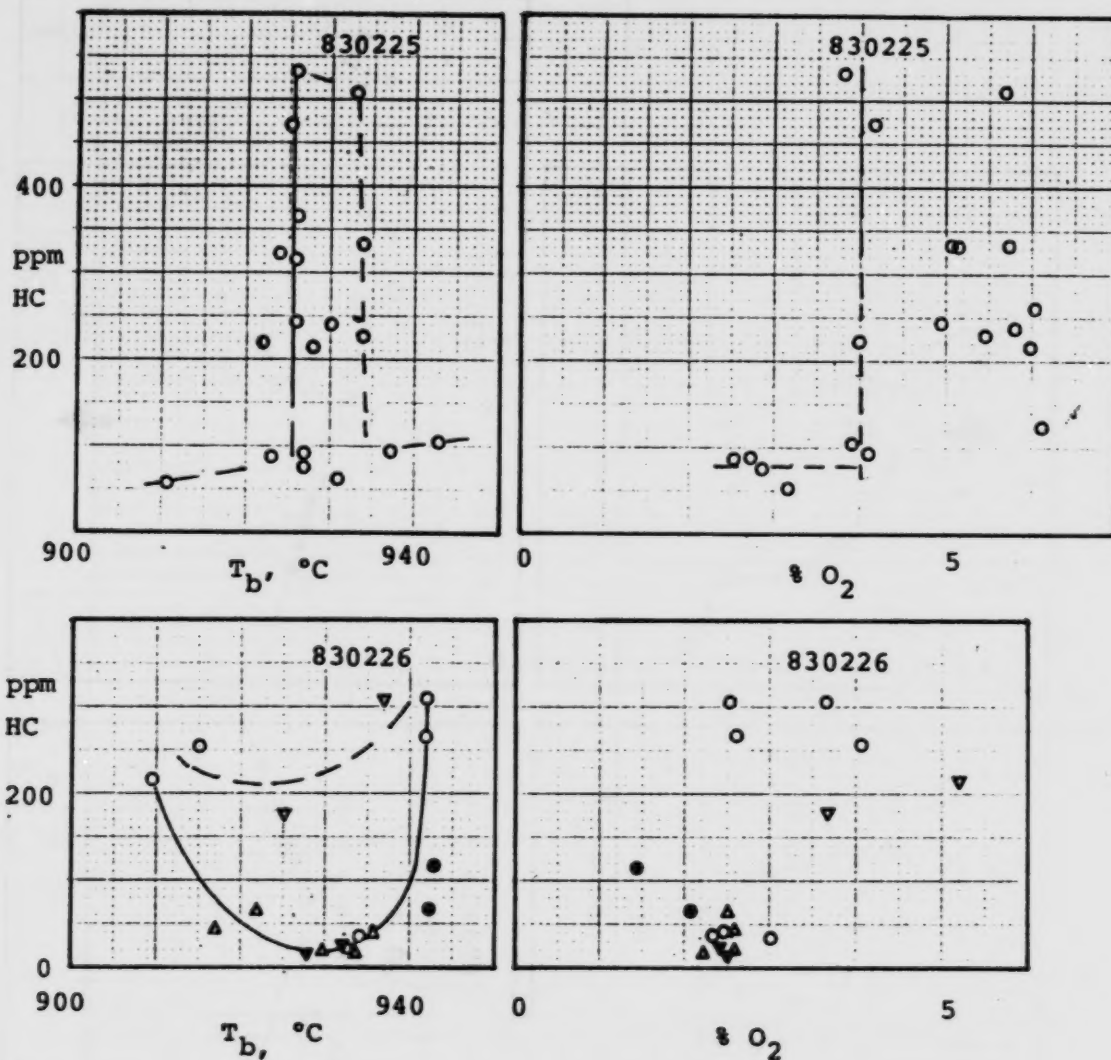


Fig. 13.9. The unburnt hydrocarbons content of the gases at combustor exit (ppm CH_4 by volume, dry basis) vs. oxygen content and bed temperature, runs 830225 and 830226. See Fig. 13.7 for key to data in run 830226.

variables {exit gas O_2 , bed temperature}, for six of the eight runs in Table 13.1. In one of the omitted runs, 830216, the fluctuation levels were all very low and the graphs like Figs. 13.1--13.8 show nothing beyond random scatter. The results on the second omitted run, 830322, are also featureless.

The results on the here interesting six runs reveal a number of significant regularities and behaviours:

1. Data shown for $t < \tau_p$ are in many cases significantly out of line, demonstrating again the magnitude of the requirement for time, measured in solids residence times, to effectively reach steady state operation.
2. At bed temperatures of 800--870 °C there was generally a tendency for exit O_2 level to drop with increasing bed temperature. Since a healthy level of excess air was supplied in all these runs, increases in coal feed (by fluctuation in the feedrate) should bring increased burning rate and increased heat release, and thus increased bed temperature and decreased exit O_2 level. The trend in exit O_2 with temperature thus agrees with expectation.
3. At the high bed temperatures of 910--965 °C, there are interesting complications in the relation between bed temperature and exit O_2 . In one case, Run 830225, there are two and possibly three bed temperature levels at given exit O_2 levels. The lower principal temperature level is a plateau at 935 °C at which exit O_2 varies from 2.5--6.2 % with no temperature change. On the upper curve, however, exit O_2 declines in a normally gradual manner with increasing bed temperature. We see here (through short segments of the statistically stationary state or overall "steady state") a possible case of the multiple steady states mentioned in Sect. 5.5. As noted there, behaviour of this sort is not unexpected when, as here, a strongly exothermic reaction is carried out in a non-adiabatic reactor. In the second case,

Run 830226, there are suggestions of a temperature plateau around 910 °C and an exit O₂ plateau around 2.5 %. The really remarkable feature, however, is the indication of strong hysteresis. The relation between exit O₂ level and bed temperature seems to follow essentially one curve when the temperature is rising or at a virtual plateau (near a maximum or minimum), but the O₂ moves to higher levels when the temperature is rapidly falling. Hysteresis produces, as seen in Fig. 13.1, a looping behaviour as bed temperature rises and falls. A certain degree of such hysteresis might be expected generally, as a natural consequence of various forms of system inertia; however, it is not discernable in the other seven runs here considered.

4. The exit CO levels show little or no trend with exit O₂ or bed temperature levels at 800--870 °C. At 910--965 °C, however, there is a clear tendency of the relation with temperature to parallel that between the exit O₂ level and temperature.

5. The results on exit SO₂ level at 800--870 °C are featureless in all runs but 830218. In the high-temperature runs, at 910--970 °C, the behaviour mirrors that of the exit O₂ levels, including hysteresis in Run 830226.

6. The most striking regularity in all the results is the virtually perfect correlation between exit NO_x level and exit O₂ level, of the simple linear form

$$X_{NO_x} = a + b X_{O_2} . \quad (13.3.1)$$

The correlation holds fast in the pre-steady-state period as well, $t < \tau_p$. The slope b is significantly higher in the high-temperature runs, at 910--970 °C. The clear practical implication of all this is that NO_x emissions can evidently be minimized by operating at the lowest excess air level that still gives acceptable combustion. The reductions in NO_x available by this device, however, are not spectacular; 20--50 % reduction is

perhaps the most that can reasonably be achieved.

7. It is interesting to see whether the correlation between NO_x and O_2 levels at combustor exit remains fixed at points upstream in the freeboard. Figure 13.8 shows that it doesn't; there is normally a shift which becomes quite marked as the splash zone is neared. Run 830329, incidentally, is the only instance of a run with a non-linear relation between the exit NO_x and O_2 levels.

8. The results on exit gas levels of HC (unburnt hydrocarbons) at 800--870 °C are featureless. At 910--970 °C, however, behaviour like that in the relation of exit O_2 level to bed temperature is evident, Fig. 13.9, including hysteresis in Run 830226.

13.4 Closure

It appears that the fluctuations in bed temperature and exit gas composition are explainable as to magnitude by attribution to variations in the coal and limestone feedrates. However, there is also evidence, particularly for temperatures above 900 °C, of multiplicity in points of steady state operation, with the possibility of oscillation between such states. As shown in Sect. 6, the combustor exit gas composition is significantly affected by freeboard reactions. This complicates matters and, although we have not attempted it here, a thorough explanation of the results will require consideration of the freeboard processes as well as those in the bed.

14. HEAT TRANSFER

14.1 Introduction

It is a characteristic of most fluidized bed combustion processes that, in order to operate at reasonable bed temperature levels, considerable heat must be taken from the bed. Adiabatic operation, which may be chosen if the object is simply to directly generate hot gas (e.g., for agricultural crop drying), necessitates the use of a great excess of air. The present work, however, was designed to fit the requirements for optimum operation of plants generating steam for electrical power generation and/or heating, and the air/fuel ratio was accordingly held close to stoichiometric, with excess air levels typically in the range 0--30 %. The adiabatic combustion temperature of dry coal under these conditions approaches 2000 °C, whereas the operating bed temperatures were in the range 800--900 °C. The resulting scale of the cooling load is roughly indicated by the temperature ratio $(2000 - 900)/(2000)$, and it was clearly substantial.

The exact magnitude of the cooling load is demonstrated in the next section. The question of greatest interest respecting the heat transfer, however, is the behaviour of the bed-side heat transfer coefficients. The combustor in the configuration used in the present work was well rigged for the study of heat transfer to in-bed tubes, but not for transfer to cooled walls (typically "water walls"), although a minor degree of wall-cooling was of necessity employed to protect metal parts at ports in the combustor wall. The behaviour of the heat transfer coefficients for the in-bed tubes is therefore examined in detail in the following sections.

14.2 The bed cooling load

The magnitude of the bed cooling load is well measured by the coefficient

$$\eta_Q = \dot{Q}_b / (-\dot{m}_f \Delta h_{c,f})$$

where $\dot{Q}_b$ is the total heat flux out of the bed, to the bed cooling surfaces, and $-\dot{m}_f \Delta h_{c,f}$ is the theoretical combustion heat release (for complete combustion of fuel to fully oxidized products, with reactants and products both at 25 °C). Values of this coefficient, Fig. 14.1, ranged from 0.26 to 0.38. As might be expected, the coefficient depends most strongly on the excess air level and the coal particle size. The results in the figure are roughly correlated by

$$\eta_Q = a - b \epsilon_a, \quad (14.2.1)$$

where ϵ_a is the excess air level expressed as a fraction, $a = \{0.32, 0.34, 0.36\}$ for the mean coal particle diameters of $\{6.6, 10.7, 19.3\}$ mm, and $b = 0.2$. The figure, of course, is presented here for illustration, not prediction. Quantitative prediction for any set of input conditions and bed temperature can be obtained from the mass and energy balances. The only difficulty is in estimating the completeness of combustion, particularly the completeness of char burnout.

14.3 Division of the bed cooling load between the cooled surfaces

As indicated in Sect. 2, heat was taken out of the bed by four routes:

- o the in-bed tubes (see description of module 3, Sect. 2.2). Of 21.3 mm outside diameter, these ran horizontally through the bed from wall to opposite wall. The exposed length of each tube (wall-to-wall) was 380 mm. Between 12 and 24 tubes were used, giving a total bed-side (outside) surface area S_o of 0.31--0.61 m². On the basis of unit bed cross-sectional area, the tube complement was $n_t/A_b = 78$ --155 tubes/m² and the bed-side heat transfer surface S_o/A_b was 2.0--4.0 m²/m².
- o the tube-port combustor wall panels (the "tube sheets" holding

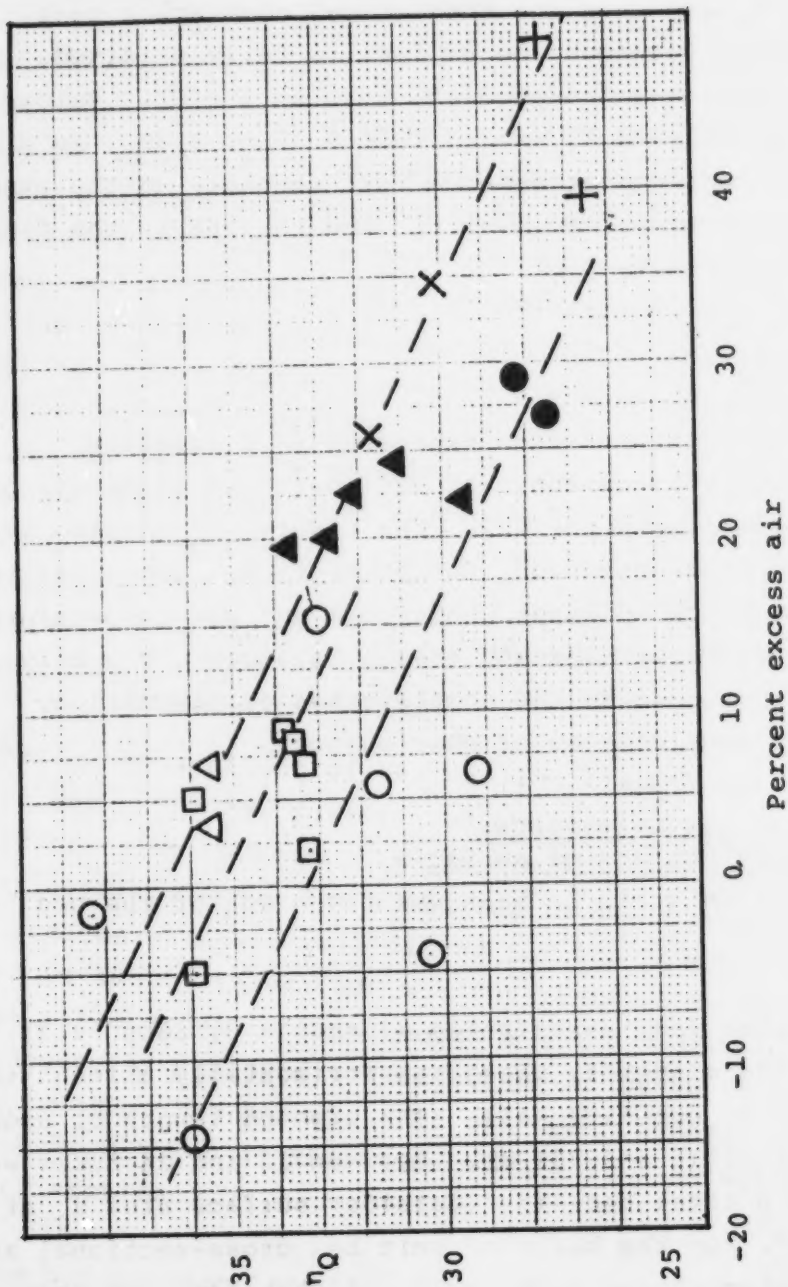


Fig. 14.1. Bed heat loss coefficient vs. excess air level. Key to data: Table N.3B.

the in-bed tubes; see description of module 3, Sect. 2.2). These were water-cooled on the outside and refractory-lined on the inside. The total nominal heat transfer surface of the two panels (including the tube ports) was 0.42 m^2 , similar in magnitude to the total in-bed tube surface. However, because of the bed-side refractory lining, these wall panels were equivalent to only about 4 in-bed tubes (or to an in-bed tube surface about $1/4$ as large).

- o the water-wall downward extension of the combustor (see description of module 2, Sect. 2.2), of 0.160 m^2 bed-side surface area. When fully active, this was equivalent to 3 or 4 in-bed tubes (or to an in-bed tube surface area $1/2$ -- $2/3$ as large). However, when coarse coal was fired, this region gradually filled with "rocks" (unfluidizably large stone particles) and thus gradually became stagnant. The heat flux simultaneously dwindled to a negligible level.
- o heat loss through the rest of the combustor wall surface. The walls here were of refractory about 140 mm thick. The nominal contact surface at 1.2 m bed depth was about 1.9 m^2 , or 3--6 times the in-bed tube surface. Because of the thick walls and low thermal conductivity, however, the heat flux by this path was negligible. A substantial part of it, directly measurable from cooling-water flow and temperatures, occurred in the neighbourhood of the one disturbance and localized sink, namely the port containing the water-cooled solids (coal and limestone) feed pipe.

14.4 Configuration and connection of the in-bed tube arrays

The geometry of the ports accommodating the in-bed tubes was discussed in Sect. 2.2 (see Fig. 2.4 and the description of module 3). Each tube was 1.12 m long and straight. The length exposed to the bed, as already noted, was 380 mm. Tubes at the same level, or at adjacent levels, were connected in series of two or four. The tube-to-tube connections were of long-radius

bend and were made of 19 mm I.D. heavy-wall rubber hose designed for service up to 90 °C. A typical array of 14 tubes in four series circuits is illustrated in Fig. 14.2. Each circuit was served by one flowmeter, and the cooling water temperature was measured at the circuit inlet and at the outlet. It was found that limiting the outlet temperature to 55--60 °C was generally sufficient to avoid boiling.

The heights of the tube ports above the distributor plate are shown in Fig. 2.4. When a circuit involved tubes at two levels, as in Fig. 14.2, the bed-side heat transfer coefficients were assumed to be associated with the mean height. Since the variation between levels was small, this procedure is reasonable. Variation with lateral position was not investigated and was assumed to be unimportant. The tubes at a given level were placed to be evenly distributed, so the circuit performance presumably represented an average for the tubes at that level.

14.5 The evaluation of heat transfer coefficients

Consider the heat transfer along a given circuit through the in-bed tubes. The differential energy balance on the cooling water over a circuit segment of differential length dx gives

$$\dot{m} c_p \frac{dT}{dx} + \dot{Q}' = 0 \quad , \quad (14.5.1)$$

where $\dot{Q}'$ is the heat flux out of the cooling water per unit length at distance x along the circuit, T is the stream-average ("bulk") temperature of the water crossing section x , and c_p is the heat capacity of the water at temperature T . The heat capacity of the water varied negligibly over the temperature range involved and was treated as a constant, 4180 J/kg K. Integration of (14.5.1) over the given circuit gives the overall energy balance

$$\dot{m} c_p (T_{out} - T_{in}) + \dot{Q} = 0 \quad , \quad (14.5.2)$$

FBC IN-BED HEAT EXCHANGER (SOUTH FACE) RUN: 821206
(YR/MO/DY)

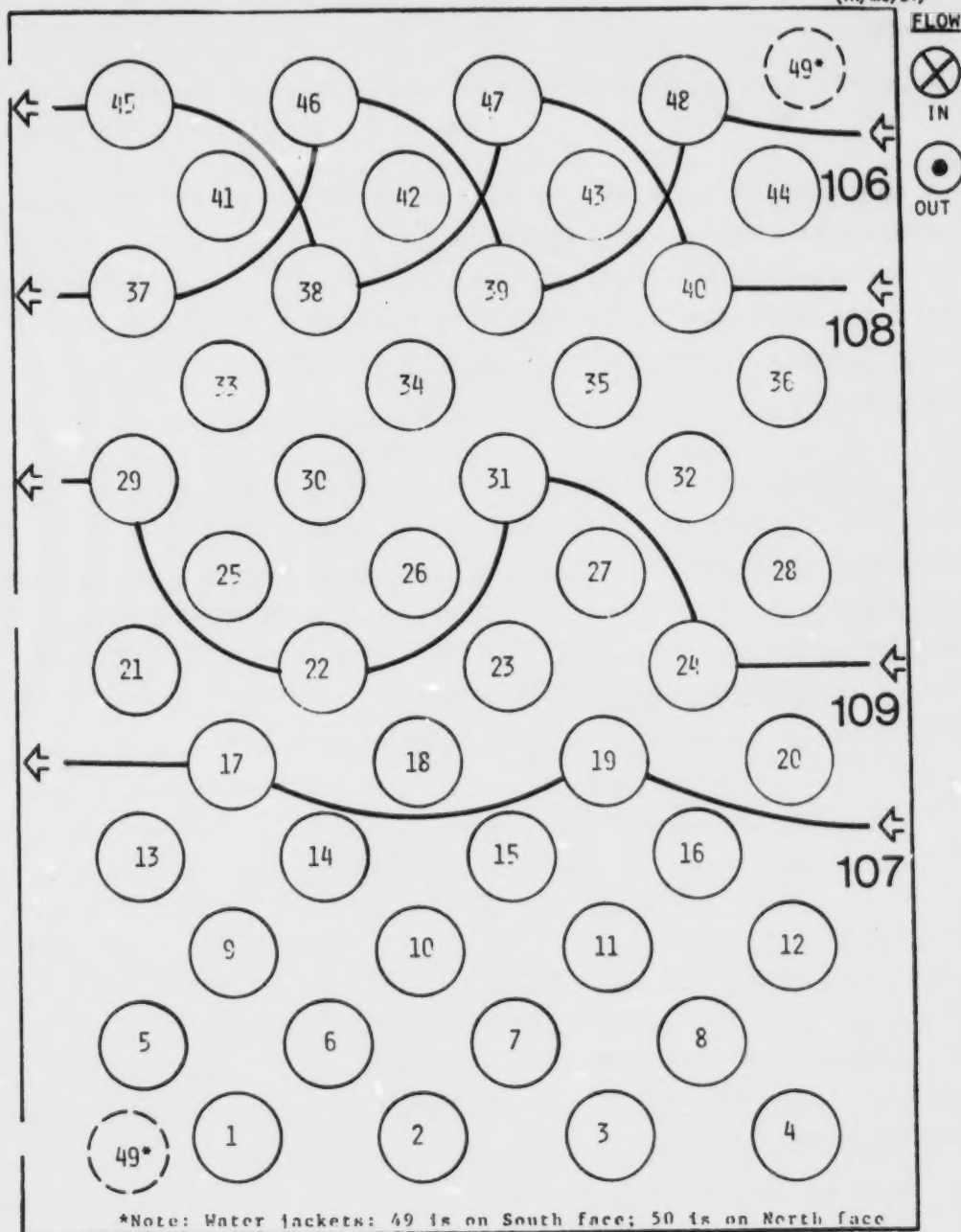


Fig. 14.2. Example of an array of 14 in-bed cooling tubes connected in four circuits, run 821206.

where

$$\dot{Q} = \int \dot{Q}' dx \quad (14.5.3)$$

is the total heat flux out of the water (negative in value, because the water was receiving heat). The circuit cooling water flowrate $\dot{m}$ and the circuit inlet and outlet temperatures were the experimentally measured variables in the study of the heat transfer, and the heat flux $\dot{Q}$ out of the circuit was thereby determined.

For the evaluation of heat transfer coefficients, certain reasonable approximations had to be made. To begin with, it was supposed that the tube circuits operated adiabatically over the portions not inside the combustor (i.e., between passes through the bed), so the heat transfer all took place through the directly exposed in-bed tube surface. Then over the exposed portions

$$\dot{Q}' = B'(\check{T} - T_b) , \quad (14.5.4)$$

where B' is the overall unit-length conductance, $W/m \cdot K$, and $\check{T} - T_b$ is the overall temperature driving force. Over the remaining portions, $\dot{Q}' = 0$. The total conductance here consists of three conductances (or resistances) in series: the tube-side or "inside" conductance, and the bed-side or "outside" conductance. According to the usual reciprocal rule for summing series conductances,

$$1/B'_i + 1/B'_w + 1/B'_o = 1/B' . \quad (14.5.5)$$

Application of the usual heat transfer laws gives

$$B'_i = \pi D_i h_i , \quad (14.5.6)$$

$$B'_w = 2\pi k_w / \ln (D_o/D_i) \quad (14.5.7)$$

$$B'_0 = \pi D_0 h_0 \quad . \quad (14.5.8)$$

The wall resistance $R'_w = 1/B'_w$ was generally negligible. Thus, with these substitutions, (14.5.1) gives

$$- \int_{T_{in}}^{T_{out}} \left(\frac{1}{\pi D_0 h_0} + \frac{1}{\pi D_i h_i} \right) \frac{dT}{T - T_b} = \frac{L}{\dot{m} c_p} \quad .$$

where L is the total in-bed length of the tubes in the circuit and $\pi D_0 L$ is the bed-side surface area. It is also implied that $T_i = T_o$, where T_i and T_o are the inside and outside tube-wall temperatures. We therefore simply write $T_w \equiv T_o = T_i$.

The bed-side heat transfer coefficient h_0 is the basic unknown variable in (14.5.9); however, it can reasonably be taken to be virtually constant along a given circuit. The tube-side coefficient h_i is function of local flow conditions and of the local temperatures T and T_w . The value of h_i may be estimated from standard correlations for heat transfer inside tubes. Very accurate determination of h_i would be difficult. As already noted, each in-bed tube was entered from a curving length of rubber hose 19 mm I.D. and more than 300 mm long. There was then a run of about 370 mm inside the $D_i = 15.8$ mm tube before the 380 mm long section inside the combustor (directly contacting the bed) was entered. This was followed by another run of 370 mm out the other side of the combustor before the next connecting section of rubber hose was entered. A fair approximation to the situation along a tube cooling circuit, ignoring the slight flow-area contractions and expansions and differences in wall roughness between hose and tube, and ignoring as well the curvature (of long radius) of the hose sections, is to say that the stream at entry into the active section of a tube was in equilibrium pipe flow, and the temperature distribution was

uniform. In principle, then, h_i ought to be treated as a function of distance along the active section of each tube, with due allowance for:

- o the temperature dependence of the fluid properties, expressed through T_w and T
- o the periodic re-establishment and growth of the thermal boundary layer.

We have critically examined the requirements for such allowances and have concluded, by comparison with rigorous computation) that the problem can be rather simply handled, with very satisfactory results, as described in the next paragraph. The reasons why such simplification is possible are:

- o the bed-side temperature difference $T_b - T_w$ varies very little along a circuit, because T_b is so large.
- o the unit-length heat flux, which can be written as

$$\dot{Q}' = \pi D_o h_o (T_w - T_b) , \quad (14.5.10)$$

varies correspondingly little.

- o the tube-side coefficient was large, typically around ten times greater than, the bed-side coefficient. Consequently it figured small in the overall heat transfer, and a modest accuracy in estimating it was sufficient.

The computation procedure finally evolved was as follows:

1. Having calculated $\dot{Q}$ for a circuit from the experimental data by means of (11.5.2), estimate

$$\dot{Q}' = \dot{Q}/L \quad (14.5.11)$$

2. Set

$$T_{w,0} = T_{in}, \quad T_{w,L} = 100^\circ\text{C}$$

where $T_{w,0}$ is the wall temperature at the start of the first active section of the circuit, and $T_{w,L}$ is the wall temperature at the end of the last active section.

3. Estimate the tube-side heat transfer coefficient h_i from the correlation of Sieder and Tate (1936; commonly cited in textbooks on heat transfer):

$$\text{Nu} = 0.027 \text{Re}^{0.8} \text{Pr}^{1/3} (\mu/\mu_w)^{0.14} \quad (14.5.12)$$

where $\text{Nu} \equiv h_i D_i / k$, $\text{Re} \equiv 4\dot{m} / \pi D_i \mu$, and $\text{Pr} \equiv c_p \mu / k$. In applying this equation, evaluate the stream fluid properties μ , c_p and k at the median stream temperature $(T_{in} + T_{out})/2$. Evaluate μ_w at the median wall temperature $(T_{w,0} + T_{w,L})/2$.

4. Estimate the exit wall temperature $T_{w,L}$ from the tube-side heat flux equation

$$\dot{Q}' = \pi D_i h_i (T - T_w) \quad (14.5.13)$$

which gives

$$T_{w,L} = T_{out} - \dot{Q}' / \pi D_i h_i \quad (14.5.14)$$

(remembering that Q' is negative).

5. Using the value of $T_{w,L}$ from step 4, repeat Steps 3--5. Iterate until the value of $T_{w,L}$ stabilizes within 1 °C or less.

6. Integration of (14.5.9) with h_i and h_o constant gives

$$h_o = \left[\frac{\pi D_o L}{\dot{m} c_p \ln \left(\frac{T_b - T_{in}}{T_b - T_{out}} \right)} - \frac{D_o}{D_i h_i} \right] \quad (14.5.15)$$

Using the final value of h_i from steps 3--5, calculate the bed-side heat transfer coefficient h_o .

Values so found for h_i were in the range 2200--4400 W/m² K, and those for h_o were 200--360 W/m² K. Thus, as indicated earlier, h_i was large relative to h_o , and its role in regulating the overall heat transfer (according to the reciprocal rule of addition of conductances in series) was correspondingly small.

The relation between the heat fluxes out of the in-bed cooling circuits, individually denoted above simply by $\dot{Q}$, and the heat flux $\dot{Q}_b$ out of the bed to those circuits, is $\dot{Q}_b = - \sum \dot{Q}$, where the sum is over all the circuits. The convention followed is the common one in transport phenomena (and its branch of heat transfer) that heat fluxes are positive when directed out of the considered system. Thus, with the cooling water as the system, $\dot{Q}$ is positive leaving the water. With the bed as the system, $\dot{Q}_b$ is positive leaving the bed.

The above argument has been kept notationally simple. In a more general discussion--e.g., in the context of a broader mathematical model--the cooling circuits would have to be explicitly identified. We might then write $\dot{Q}_j$ for the heat flux out of the j 'th cooling circuit, $\dot{m}_j$ for the cooling-water mass flux in the circuit, $h_{o,j}$ for the bed-side heat transfer coefficient, etc. The labels here generally represented by j might be taken as the stream labels assigned to the cooling-water streams (one to each stream).

14.6 Vertical variation of the bed-side heat transfer coefficient

Figure 14.3 shows the typical variation of the bed-side heat

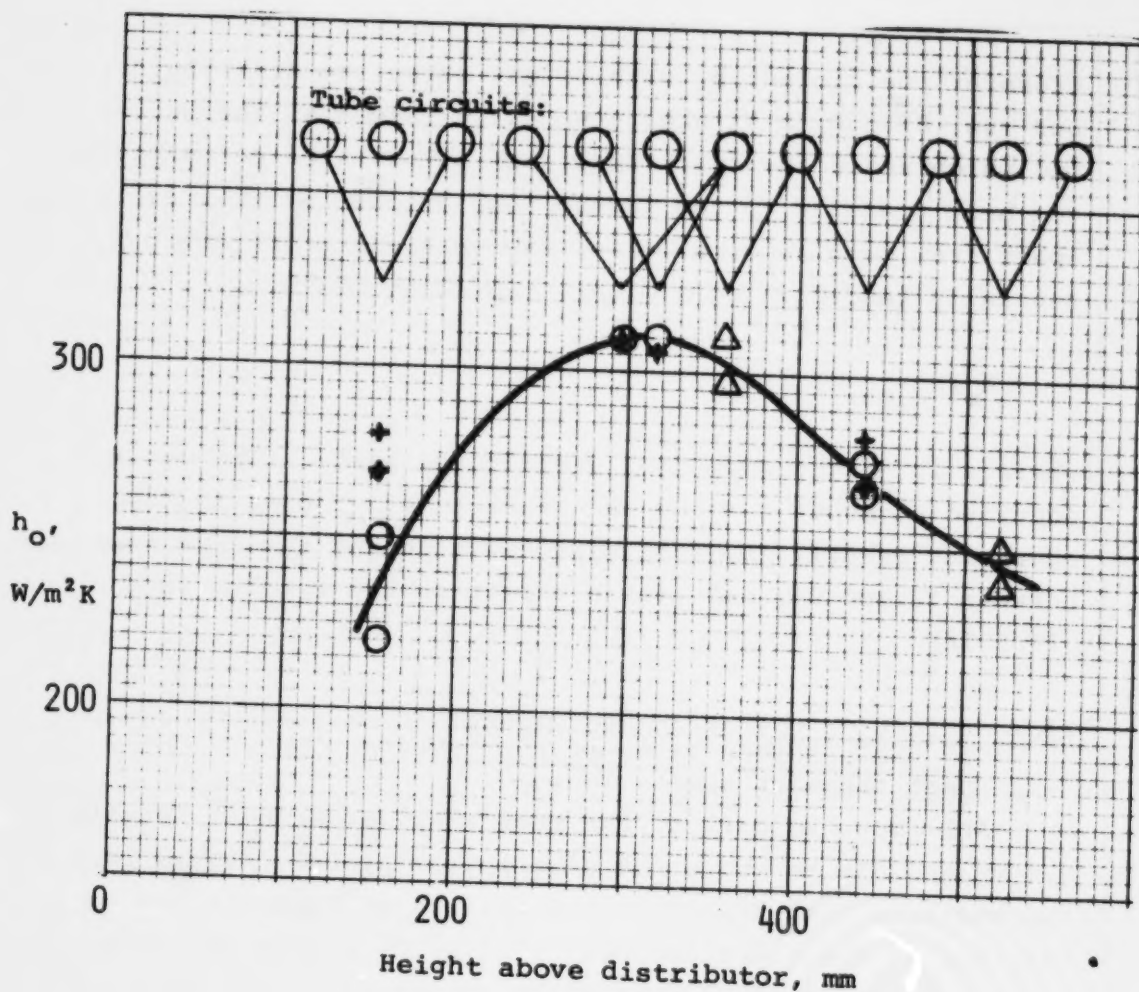


Fig. 14.3. Variation of the bed-side heat transfer coefficient for in-bed tubes with height of the tubes (or the mean height of a multiple-tube circuit) above the distributor. Runs: $\circ$ +, 821130; Δ , 830215. Operating time in 821130, hr: $\circ$, 6; +, 8, clinker forming.

transfer coefficient h_o with height above the distributor. In one run, 821130, a clinker eventually formed at the bottom of the bed which ended the run; the figure illustrates perturbation of the heat transfer to the lowest tube circuit by this disturbance. Apart from such abnormalities, the vertical profiles of h_o generally tended to be similar, representable by

$$h_o/h_{o,max} = f(z) , \quad (14.5.16)$$

where z is height above the distributor plate. The maximum occurred at about $z = 300$ mm. Effects of changes in the operating conditions, including variation of the tube complement between 12 and 24, had little effect on the results in the form (14.5.16).

14.7 Effect of operating conditions on the absolute value of the bed-side heat transfer coefficient

All runs had part of the tube array at or near heights of 488 mm and 650 mm above the distributor plate. Figure 14.4 shows the bed-side heat transfer coefficients at these levels as a function of the elutriant dust ("flyash") efflux per unit of bed cross-section, $\dot{m}_e/A_b$. Statistical analysis of the data indicates a weak dependence on the mean limestone particle diameter and the fluidizing velocity. It would take a much larger set of data to determine the form of this dependence with some exactitude. However that might be, assumption of a power-law dependence,

$$h_o = C v^a \hat{D}_p^b , \quad (14.7.1)$$

gives a good correlation, Table 14.1 and Fig. 14.5. It seemed to us at one time that the elutriation dust flux, a measure of the flux of fine particles rapidly passing up the bed, has a measurable influence (tending to increase the heat transfer coefficient), but close statistical examination has indicated that the effect is negligible. Coal particle diameter, too, has

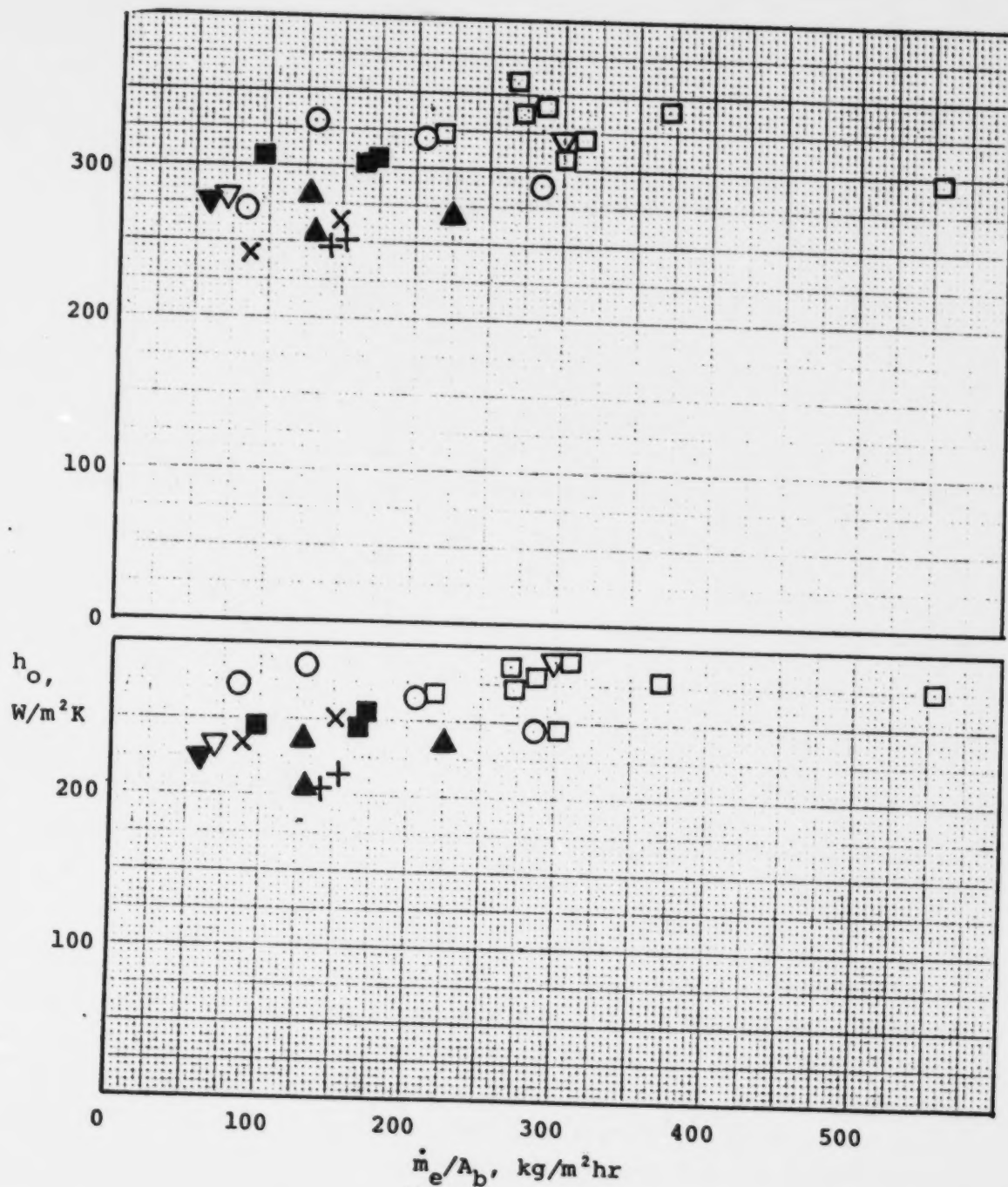


Fig. 14.4. Bed-side heat transfer coefficient for tubes vs. the elutriated solids flux density. Height of tube circuits above distributor, mm: top graph, 488; bottom graph, 650. Limestone particle diameters $\hat{D}_p$, mm: $\circ$, 0.49; $\square$, 0.78; $\triangle$, 1.72; ∇ , 1.98. Fluidizing velocities v , m/s: $\times$, 1.6--2.2; $+$, 2.7--3.0.

Table 14.1. Statistically fitted exponents and constant in the correlation for the bed-side heat transfer coefficient,

$$h_o = C v^a \hat{D}_p^b$$

Coal	z, mm	a	b	-a = 0.2, -b = 0.1	
				C	s _C
				(W/m ² K) (m ³ /s ³) ^{0.1}	
Minto	488	-0.180	-0.099	147	7
	6.50	-0.196	-0.103	176	10
Devco	488	--	--	157	8
	6.50	--	--	139	8

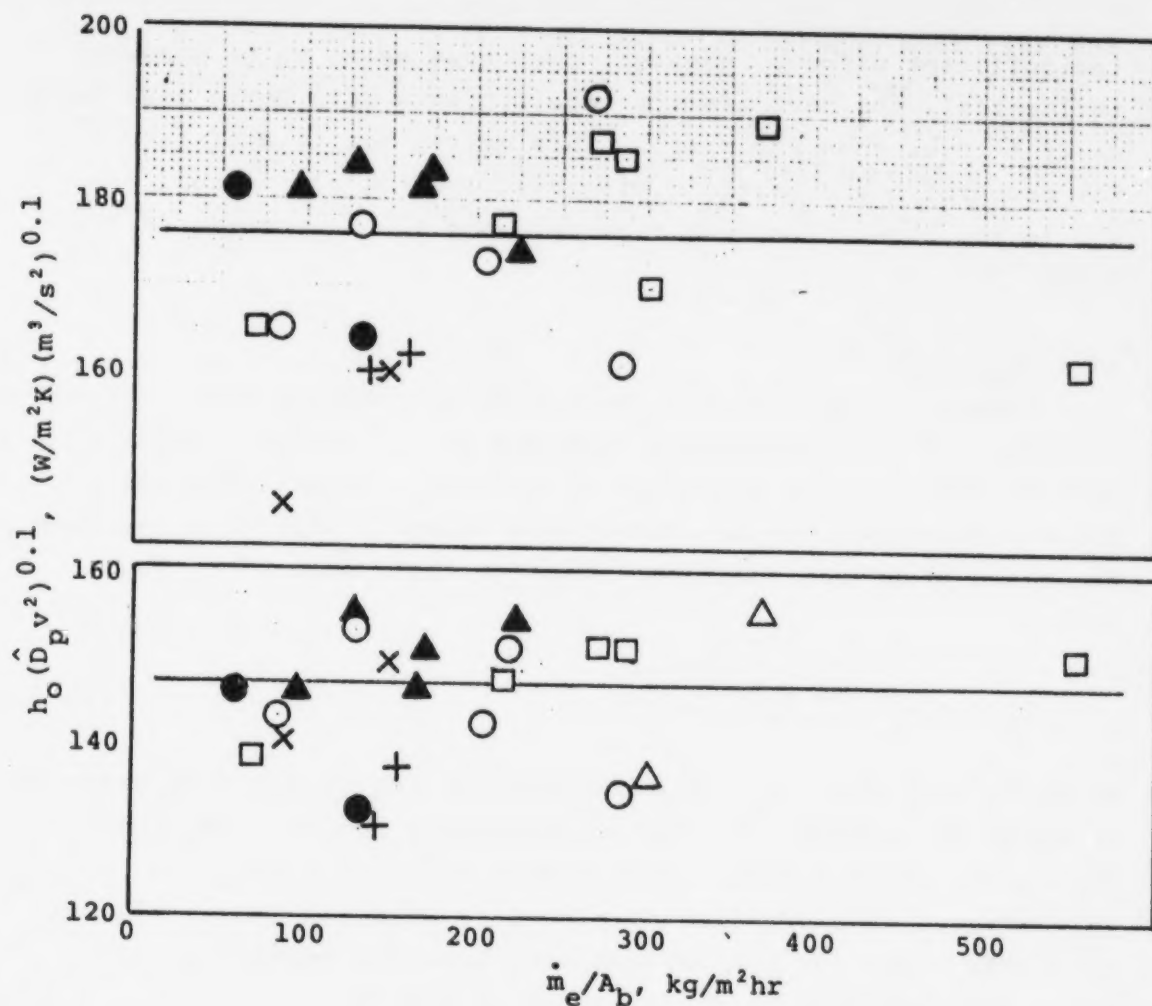


Fig. 14.5. Correlation of bed-side heat transfer coefficient for in-bed tubes with limestone particle diameter and fluidizing velocity. Height of tube circuits above the distributor, mm: top graph, 488; bottom graph, 650. Ordinate: point-by-point values of $C = h_o(v^2 \hat{D}_p)^{0.1}, (\text{W/m}^2\text{K})(\text{m}^3/\text{s}^2)^{0.1}$. The horizontal lines are the statistical means for the Minto trials. Key to data: Table N.3B.

insignificant effect. However, coal type seems to be somewhat influential, the heat transfer coefficients with Devco coal being 8--20 % lower than for Minto. This may be due to differences in the proportion and particle size distribution of the coal ash, and in the balance between freeboard and in-bed burning of the coal.

14.8 Closure

Exhaustive studies have been made of bed-side heat transfer coefficients on single tubes immersed in a fluidized bed of uniform non-reacting particles at ordinary temperatures, well below combustion levels. Grewal and Saxena (1980) have reviewed these and presented additional data of their own. They found that

$$h_o = v D_o^{-1/3} D_p^{-1}, \quad (14.8.1)$$

which is very different from the present result for heat transfer to tubes in bundles, in beds of relatively coarse, reacting particles, at high temperature levels and high fluidizing velocities. An expression developed by Gelperin (Botterill, et al. 1980) gives for the maximum heat transfer coefficient in a fully immersed tube bundle in a bed at 850 °C

$$h_o = 34 D_p^{-0.34}. \quad (14.8.2)$$

For average limestone particle diameters of 0.0005--0.002 m, the range in our work, this gives $h_o = 280\text{--}450 \text{ W/m}^2 \text{ K}$, nearly bracketing the range of our experimental maxima of h_o , 240--360 W/m^2 . There is here a fair measure of agreement on the value of h_o , at least, if not on the exponent for D_p .

Among the other units of heat exchange equipment associated with the pilot plant, besides bed-cooling, the only one worth mentioning--because it is directly influenced by combustor performance--is the freeboard heat exchanger (see description of

modules 8 and 9, Sect. 2.2). The relation between heat flux and cooling water flowrate for this unit, Fig. 14.6, was virtually linear and was little or negligibly influenced by other operating conditions. Mathematical modelling of the performance of this heat exchanger is an interesting exercise, but outside the present scope.

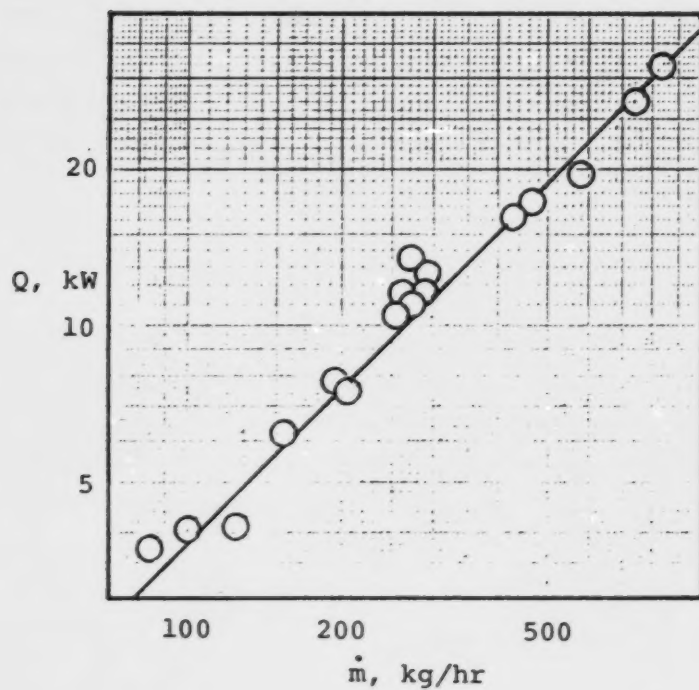


Fig. 14.6. Performance of the freeboard heat exchanger: heat flux vs. cooling water flow. Data from all runs.

15. ASPECTS OF FLUIDIZATION DYNAMICS AND EFFECTS THEREOF

15.1 Introduction

As noted earlier, the bed depth was regulated in the present work by the overflow port, 1.19 m above the distributor, from which overflowing bed material was collected and removed as "bed product". This does not mean that the expanded bed depth was exactly 1.19 m. It appears, rather, that the bed stabilized with the overflow port somewhere in the splash zone, and possibly relatively high in it rather than low.

Bed characteristics that are of direct significance for performance in fluidized-bed combustion are the solids holdup m_b and the minimum fluidizing velocity v_{mf} . The holdup per unit cross-sectional area and the pressure drop across the bed are closely related. In the technique of controlling bed depth (or holdup) by forcible extraction of bed material rather than by gravity overflow, pressure drop is usually used as the actuating signal, and extraction rate is the manipulated variable. In some of our planned future work, underbed extraction in this manner will be used to provide substantially greater bed depths, greater holdups, and correspondingly greater solids residence times than are possible with the present single, fixed overflow port.

15.2 Bed pressure drop

The bed pressure drop was measured directly as the pressure difference between a pressure tap at the bottom of the bed (in the wall, just above the distributor) and one located in the freeboard.

According to simplest theory, a fluidized bed is fully levitated by the drag of the rising gas stream, so

$$(-\Delta P)_b A_b = g m_b (\rho_p - \rho_g) / \rho_p + g \rho_g A_b \Delta Z \quad , \quad (15.2.1)$$

where $(-\Delta P)_b$ is the static pressure difference measured as just described, Δz is the difference in elevation between the pressure taps, ρ_p is the particle density, ρ_g is the gas density, and g is the acceleration due to gravity. Since $\rho_g \ll \rho_p$, this reduces to

$$(-\Delta P)_b A_b = g m_b , \quad (15.2.2)$$

where $(-\Delta P)_b$ is the pressure drop across the bed, virtually indistinguishable from $(-\Delta P)$ in (15.2.1). A close relationship is thus expected between the bed mass or "solids holdup" m_b and the bed pressure drop. To make the relation exact, it is written in the useful working form

$$(-\Delta P)_b A_b = K_b g m_b , \quad (15.2.3)$$

where the coefficient K_b is expected to be near unity. Experimental values of K_b are presented in the next section, as part of the discussion of bed holdup.

Since bed holdup in the present work was regulated by bed overflow from the fixed overflow port, it varied with the bed expansion and splash characteristics. These factors in turn varied with the operating conditions. The results, Fig. 5.1, indicate that fluidizing velocity and limestone particle size were most influential. Increasing the fluidizing velocity certainly increased the bed expansion and the relative extent of the splash zone and thus, as expected decreased the holdup and the bed pressure drop. Bed pressure drop was generally in the range 400--560 mm w.c. (water column), or 400-560 kg_f/m², or 3900--5500 Pa.

15.3 Bed Holdup

15.3.1 General remarks

As mentioned earlier, solids holdup in the bed was measured

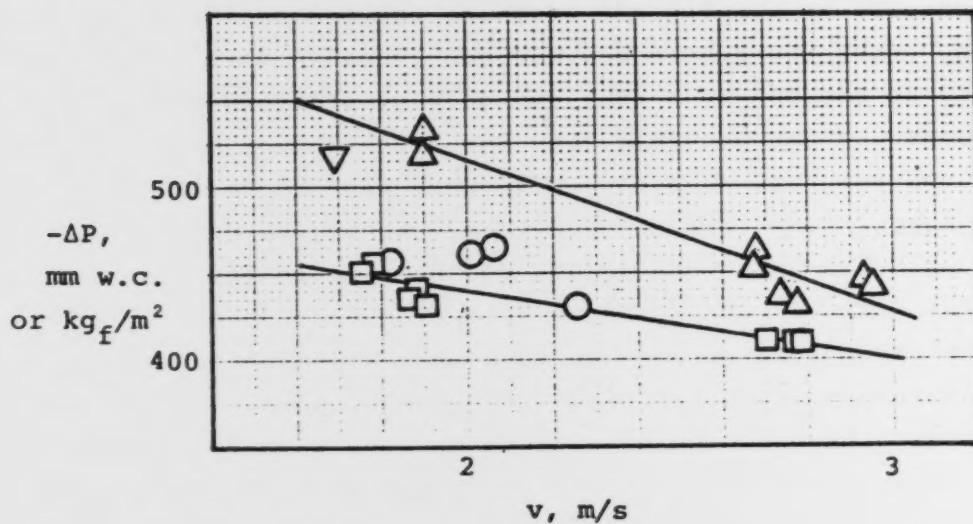


Fig. 15.1. Bed pressure drop vs. fluidizing velocity, for runs with the water-wall extension in place. Limestone particle diameter $\hat{D}_p$, mm: ○, 0.49; □, 0.78; △, 1.72; ▽, 1.98.

experimentally by dumping and weighing the bed solids at the end of a run, after the system had cooled. Because this could be done only at the end of a run, and only by shutting down, the measurement was not possible when the next run was begun without pause. The results, when obtained, were also of variable significance. Some runs ended with clinkering--a catastrophe--and the material then left in the bed bore no relation to what was there in smooth steady-state operation. Occasionally runs had to be terminated because of equipment problems before steady-state operation had been fully established, but not before some useful data had been obtained. In runs with the coarse (19.3 mm dia) Minto coal, coarse stony matter occurring in the coal accumulated at the bottom, forming a stagnant zone. This material appeared to be nearly all shale, consisting of characteristically flattish, elongated particles, in size up to 50 mm long.

15.3.2 Holdup in normal steady-state operation

In the absence of significant accumulation of unfluidizable stony matter entering with the coal (thus in all runs except some of those with the 19.3 mm Minto coal), the bed holdup m_b was 70--90 kg or, on the basis of unit base area, $m_b/A_b = 450\text{--}580$ kg/m². The coefficient K_b in the relation (15.2.3) with the bed pressure drop was about 0.90, and this value of K_b was used to estimate the holdup from the bed pressure drop in those runs where the holdup was not measured directly.

15.3.3 Holdup in runs with bed stratification

It was found in the runs with the coarse (19.3 mm) Minto coal that stony matter held on a 4 mm standard (5 mesh) sieve was virtually absent in the bed overflow. This material was, however, prominently present at the bottom of the bed and obviously formed the stagnant zone which, as it built up, progressively blocked heat transfer to the water-wall downward extension of the combustor (the 100 mm high first module just

above the distributor plate). Operation was never continued much beyond the point of filling the water-wall extension with stagnant material, since the underbed injection of coal and limestone would have been interfered with. The maximum accumulation of + 4 mm sieve-size stone material was around 12 kg. The remaining mass of the bed, composed of - 4 mm material, was of the normal holdup size. The bed pressure drop was normal as well. Thus the retention of a modest amount of unfluidizable material seemed to have scant effect on the holdup of fluidizable (non-stagnant, circulating) material and on the pressure drop.

In calculating solids residence times here, $\tau_p = m_b / \dot{m}_p$ and $\tau_b = m_b / \dot{m}_b$, the stagnant material was--appropriately-- ignored, and the effective bed mass m_b was estimated in the usual manner (with $K_b = 0.9$) from the bed pressure drop.

15.4 Bed depth

The bulk density of the bed product was around 1150 kg/m³. For bed holdups in the observed range 450--580 kg/m², this indicates rest bed-depths of 400--500 mm. The single sapphire window in the side of the combustor which gave a view into the bed was quite inadequate for visual observation of the expanded bed depth. It is intended in future work to provide pressure taps at closer intervals than in the present system to allow determination of the effective expanded bed depth from the vertical profile of the static pressure.

15.5 The minimum fluidizing velocity

The minimum fluidizing velocity, v_{mf} , was measured in a number of runs, always at the end of the run, because of the disturbance it caused. The technique was to shut off the cyclone recycle air and then reduce the flow of fluidizing air by steps, while observing the bed pressure drop at each step. This procedure had to be carried out quickly, before the bed temperature, or any other significant bed characteristic, changed appreciably. The procedure was not always successful because the

high reduction of flowrate required often brought the manometer reading for the orifice flowmeter below the level that could be accurately observed. The addition of a wide-range differential pressure transducer would solve this problem, and is planned for future work.

A potential pitfall should be noted as well. When using material with a range of particle sizes, the larger sizes may settle down and defluidize as the gas flow is reduced. This problem was certainly present to some degree in runs where the 19.3 mm Minto coal was burned.

Having obtained data on bed pressure drop as a function of air flowrate as just described, a log-log graph of the results was prepared. The data in the fluidization region, as illustrated by Fig. 15.2, were fitted by a nearly horizontal straight line. Those in the fixed-bed region were fitted by another straight line with a slope of, or very near, 45°. The point of intersection of these lines was taken to define the point of minimum fluidization. Because the operating conditions in respect to gas composition were upset during its determination, the minimum fluidizing velocity was simply calculated as the superficial air velocity (with O₂ unreacted) at bed temperature and freeboard pressure:

$$v_{mf} = \dot{n}_a \tilde{V}^* (T_b/T^*) (P^*/P_{fb}) \quad (15.5.1)$$

The fluidizing velocity in normal steady-state operation, however, was based on the actual molar gas flow at combustor exit:

$$v = \dot{n}_g \tilde{V}^* (T_b/T^*) (P^*/P_{fb}) \quad (15.5.2)$$

Values of the minimum fluidizing velocity v_{mf} obtained when firing the 6.6 mm Minto coal with the 0.78 mm and 1.72 limestones are shown in Table 15.1. Definitive measurements were not

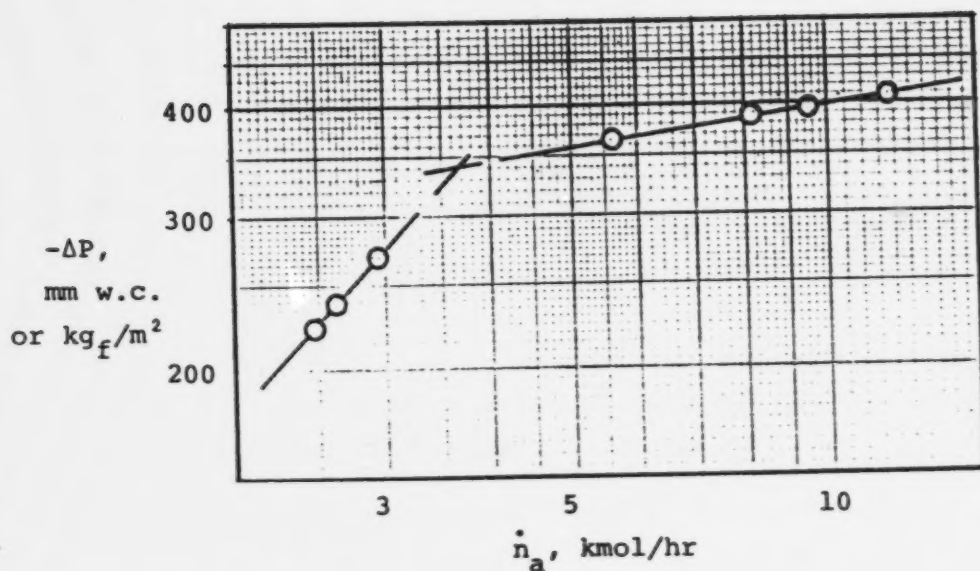


Fig. 15.2. Example of a determination of the minimum fluidizing velocity: bed pressure drop vs. air flowrate, run 830226.

Table 15.1. Experimental values of the minimum fluidizing velocity. Coal: Minto, 6.6 mm mean dia. Limestone: mean particle diameters in table.

Run	$T_b,$ °C	$\hat{D}_p,$ mm	$v_{mf},$ m/s
830201	823	0.78	0.237
830226	918	1.72	0.64

obtained in runs with the 0.49 mm and 1.98 mm limestones.

Broadhurst and Becker (1975) developed general correlations for predicting the minimum fluidizing velocity and the minimum slugging velocity. When the particle Reynolds number is not too high, the equation for v_{mf} reduces to

$$v_{mf} = 0.0023 (gD_p)^{1/2} (g\rho_p D_p^3/\mu^2)^{0.425} \quad (15.5.3)$$

The particle density ρ_p of the present bed material was roughly 2000 kg/m³. For the gas velocity μ , take the viscosity of air; over the present bed temperature range,

$$\mu = 5.62 \times 10^{-7} T^{0.623} \quad , \quad (15.5.4)$$

kg/m s. With these substitutions (15.5.3) yields the working formula

$$v_{mf} = 11.8 D_p^{1.775} T_b^{-0.53} \quad , \quad (15.5.5)$$

where D_p is in millimetres, T_b in kelvins, and v_{mf} in m/s. Assume that the appropriate mean particle diameters of the present bed materials are reasonably well represented by the mean limestone particle diameter $\hat{D}_p$. Table 15.2 shows the values of v_{mf} at 850 °C bed temperature predicted by (15.5.5) for our four limestone sizes, and also the experimental values from Table 15.1 adjusted to 850 °C. The agreement is remarkably close in view of the fact that the present bed material was of mixed particle size whereas the correlation of Broadhurst and Becker (1975) is based on data for uniform particles.

The maximum mean particle Reynolds number at the point of minimum fluidization, $Re_{mf} = v_{mf} \hat{D}_p / \nu$, was 13.4. The maximum value of the parameter $g(\rho_p - \rho_g) \hat{D}_p^3 / \rho_g \nu^2$ was 2.4×10^4 . These values are low enough, but only just, so that the flow regime in the bed at minimum fluidization was effectively a creeping flow. This

Table 15.2. Experimental and predicted values of the minimum fluidizing velocity at 850 °C bed temperature. Predicted values from (15.5.5), based on the correlation of Broadhurst & Becker (1975). Particle diameter is taken to be the mean limestone particle diameter, and the particle density as 2000 kg/m³.

$\hat{D}_p$, mm	v_{mf} , m/s, experimental	v_{mf} , m/s, predicted
0.49	--	0.081
0.78	0.231	0.184
1.72	0.68	0.80
1.98	--	0.96

existence of creeping flow has a number of implications from the theory of such flow in particle beds:

1. It confirms the validity of (15.5.3) for the present range of operating conditions.

2. It indicates that the local gas flow in the "emulsion" (the dense phase of the aggregatively fluidized bed, to be distinguished from the dilute, typically "bubble", phase) was effectively a creeping flow under all conditions in our work.

3. It indicates that when the gas flow is reduced below the point of minimum fluidization and the bed subsides to a fixed-bed state, the bed pressure drop (at effectively constant temperature and gas composition) should become directly proportional to the gas flowrate, as indeed appears to be the case in Fig. 15.2.

4. It indicates that the correct mean particle diameter to use in (15.5.3) and (15.5.5) is the hydraulic mean diameter

$$\langle D_p \rangle_h = 6 \sum V_p / \sum S_p = 6 V / S \quad (15.5.6)$$

where V is the total volume of the particles in a large sample and s is the surface area. This differs from the mean particle diameter here employed--the mass-median diameter based on sieve analysis:

$$\hat{D}_p = (\text{diameter of sieve opening that passes } 50\% \text{ by mass of the particles in a large sample}). \quad (15.5.7)$$

However, with the possible exception of the 0.49 mm limestone, the particle size distributions in our limestones were quite narrow, so this factor was likely not very important for the application of (15.5.3) and (15.5.5). In any case, a practical viewpoint is that since (15.5.5) seems to give good predictions of v_{mf} for operation under conditions like the present ones, with

D_p defined on the limestone particle size as just described, this equation can be regarded as a satisfactory correlation of the present results and a valid relation for predicting v_{mf} on the basis of limestone particle size.

It may be noted that if V in (15.5.6) is taken to be the volume of the particles in unit (bulk) volume of a sample, then $V = 1 - \epsilon$ and $S = S'''$ so

$$\langle D_p \rangle_h = 6 (1 - \epsilon) / S''' \quad (15.5.7)$$

where ϵ is the "fraction void" (the fractional void volume) and S''' is particle surface area per unit bulk volume.

Since the operating range of the fluidizing velocity was 1.6--2.9 m/s, the results in Tables 15.1 and 15.2 show that the ratio v/v_{mf} covered the very broad range 1.7--36.

15.6 Quality of fluidization

Since the bed could not be observed visually to any significant extent, and studies were not made using other diagnostics such as pressure fluctuations, it is difficult to say much about the quality of fluidization. Some indication is found, though, by examining the tendency for slugging. The correlations of Broadhurst and Becker (1975) give for the minimum slugging velocity at $1 < H_s/D < 3$ in beds of round cross-section.

$$v_{ms} = 5.1 (D_b/H_s)^{0.895} (gD_p)^{0.5}, \quad (15.6.1)$$

where D_b is the bed diameter and H_s is the "settled" or rest bed depth. Since the present beds were of square cross-section, the effective bed diameter is suitably defined as $D_b = (4 A_b/\pi)^{1/2} = 0.444$ m. Setting $H_s = 0.5$ m, the approximate present value, (15.6.1) gives

$$v_{ms} = 0.46 \sqrt{D_p}, \quad (15.6.2)$$

where D_p in millimetres and v_{ms} in m/s. A graph can then be drawn, Fig. 15.3, showing the minimum slugging velocity v_{ms} calculated from this formula and the minimum fluidizing velocity v_{mf} from (15.5.5) as functions of particle diameter D_p . It is predicted that $v_{ms} < v_{mf}$ at $D_p > 1.43$ mm, which is impossible. What the graph actually suggests is that for our 1.72 mm and 1.98 mm particle sizes the regime of smooth aggregative (bubbling-bed) fluidization was virtually non-existent, the slugging tendency being present from the beginning. Further, v_{ms} was generally considerably smaller than our operating values of the fluidizing velocity v , so our beds were all operated in or beyond the slugging regime. The state beyond is that of turbulent fluidization.

A slugging bed generally exhibits relatively poor solids mixing. The presence of the in-bed tube bundle, with tubes uniformly spaced laterally, should however be ameliorative, tending to break up slugs and promote dispersion.

15.7 Closure

There is some confusion in the field about the definition of the operating bed depth in fluidized bed combustion. What is meant when someone says, "We operated at a bed depth of 1 m"? Is it the rest bed depth, the bed depth at incipient fluidization, the expanded bed depth, or what? In the present work the nominal bed depth, defined by the location of the overflow port, was 1.19 m. The overflow port, however, was certainly fairly high up in the splash zone, so the expanded bed depth--if such a concept can be defined for a violently agitated bed at $v > v_{ms}$, was likely significantly less, perhaps closer to the rest depth of approximately 0.5 m than to the maximum possible level of 1.19 m. In any case, this question and others related to the general problem of the quality of fluidization are being more closely investigated by means of a cold model of our combustor, of about 1/3 the cross-sectional area, operated with cold air and with

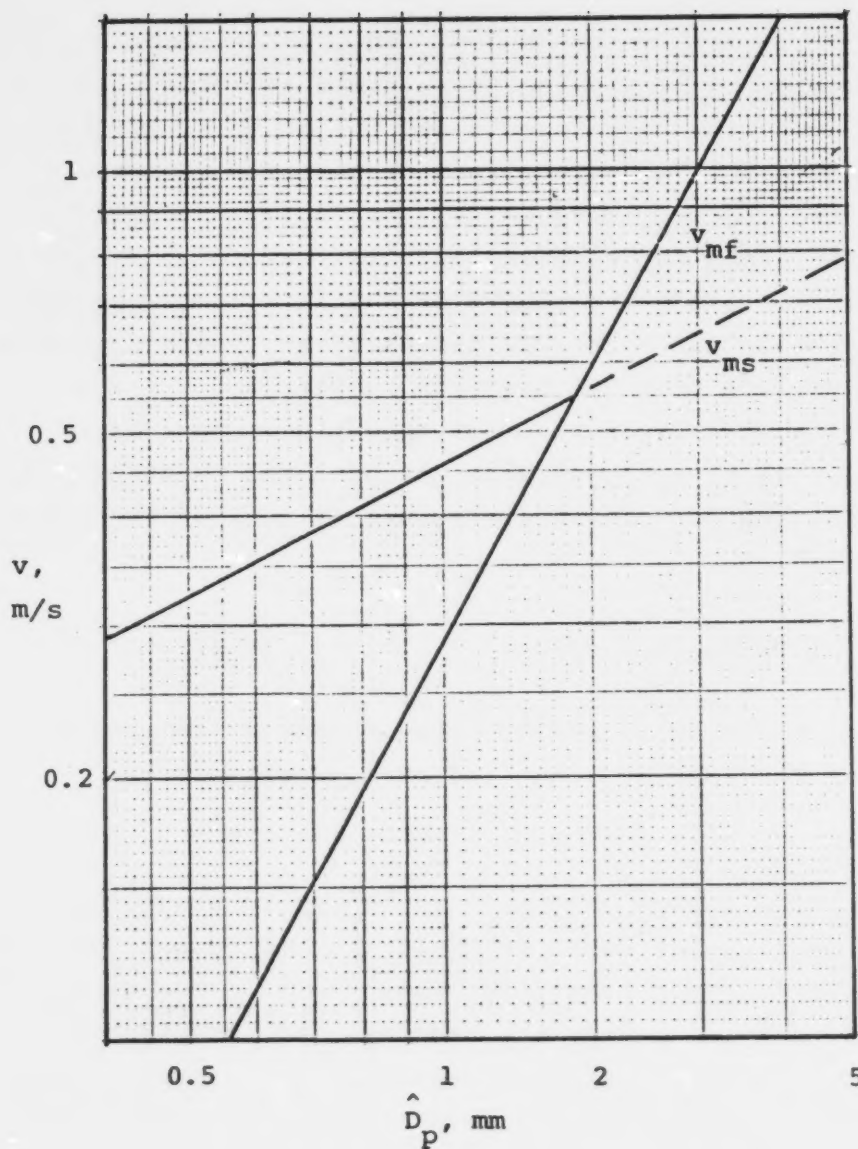


Fig. 15.3. Minimum fluidizing velocity and minimum slugging velocity as functions of particle diameter at 850 °C. Computed from the correlations of Broadhurst and Becker (1975).

spent bed material from the pilot-plant combustion runs described herein. This work is part of the Ph.D. thesis research of Mr. Fernando Preto.

It does appear that our "approximately 1 m deep" beds were in substance shallower than those of most investigators and practitioners, having perhaps $1/2$ -- $2/3$ as much holdup per unit area (or $1/2$ -- $2/3$ times the rest bed depth). We have plans for future work in which the effect of "bed depth" will be determined, using holdups up to twice the present (rest bed depths up to 1 m). This will increase the solids residence time proportionately, and similarly the time required to effectively reach steady state operation.

16. OPERATING AND ANALYTICAL PROBLEMS

16.1 Introduction

In the early phases of the work, nearly all the problems were connected with solids feeding and removal. Components of the cyclone ash recycle line rapidly wore out. The air-lock valve in the coal/limestone feedline frequently jammed. The water-cooled auger removing the bed product once jammed, filled with hot solids. These and many other operating problems had to be solved, and were, and then ceased to be matters of concern. Problems in the functioning of the instrumentation then took major importance, and these too were dealt with. The following two sections concern some significant problems for which fully satisfactory solutions have not yet been found, or which are potentially of an abiding nature--always threatening but perhaps avoidable.

16.2 Clinkering

There were some early runs, mentioned in Sect. 1, in which serious clinkering took place. It did not occur in later ones because the conditions favouring it were recognized and avoided. What we wish to do here is to record our understanding of those conditions.

The clinkers were generally formed in the tube bundle and grew until they forced stoppage of operation. It was observed that this happened only when firing the coal of largest diameter (the 19.3 mm Minto) with the limestone of largest diameter (the 1.98 mm size) at the low level of fluidizing velocity (1.6--1.9 m/s). It appears that the problem arose primarily because of poor quality of fluidization and even worse solids mixing. The large lumps of coal would naturally tend to circulate through the cooling-tube bundle less freely than small particles. The fluidizing velocity near the distributor plate, where air entered cold, was likely locally inadequate to provide more than a very

feeble agitation low down in the bed. The use of underbed feeding in these circumstances, with coal and limestone injection low in the bed, was probably unhelpful too. All these factors likely combined to create local pockets of stagnation and overheating. The clinkers were not particularly carbonaceous but seemed to be made up of average bed material all fused together.

16.3 Reliability of solids analyses

Indications of the reliability of the solids analyses have already been given in discussing the various element balances. A further check on the accuracy of the major analyses (S, C, H and N) was made by having samples of the same material analyzed by different laboratories. Our own sulphur analyses, Table 16.1, agree well with determinations at the Energy Research Laboratories, CANMET. We presently have no equipment for C/H/N analysis, so we relied entirely on a commercial analyst who had been recommended to us, analyst "A" in Table 16.1. Checks were run by two other commercial analysts, "B" and "C", and at the aforementioned CANMET laboratory. The table shows reasonably good agreement on the hydrogen content of coal. The agreement on everything else is spotty and not what we expected. In particular, the C and N analyses for coal are all significantly different. It appears that the problem is one of instrumental analysis. Some instruments, though evidently good for easy materials such as most organic chemicals, pharmaceuticals, and foodstuffs, can be difficult to manage in work on relatively refractory and unreactive materials such as coal, coke and char. In any case, thought and care should always be given to optimum sample preparation and handling. There are two models of instruments now available which seem to have all the features necessary to do a good job. We believe the best solution for our problems is to get one of these instruments and train our technicians in its use with our materials. This should give the best quality of analysis, tailored to our requirements. Funds are currently being sought for this purpose.

Table 16.1. Comparison of solids analysis by various laboratories.

Stream	Run	Sample	Analyst	Weight %, dry basis					
				H ₂ O	Ash	S	C	H	N
Minto Coal	—	821026	Queen's	0.77	20.05	7.39	—	—	—
			EMR-ERL	0.80	20.52	7.54	66.4	4.14	1.19
			Comm. A	—	—	—	71.4	4.27	0.73
			Comm. B	—	—	—	68.3	3.98	0.54
			Comm. C	—	—	—	66.6	4.35	0.83
Devco Coal	—	830310	Queen's	2.37	23.22	5.36	—	—	—
			EMR-ERL	1.75	23.20	5.61	64.3	4.46	1.50
			Comm. A	—	—	—	67.6	4.08	1.01
			Comm. C	—	—	—	59.4	4.13	1.13
Bed product	830201	1400-1600	Queen's	0.00	96.07	9.90	—	—	—
			EMR-ERL	0.00	94.51	9.66	2.82	0.12	0.22
			Comm. A	—	—	—	2.34	0.33	0.20
			—	—	—	—	2.16	0.37	0.07
			Comm. B	—	—	—	4.32	0.20	0.29
			Comm. C	—	—	—	2.45	0.58	0.02
			—	—	—	—	—	—	—
Cyclone	830201	0525	Queen's	0.23	70.63	5.02	—	—	—
			EMR-ERL	0.34	69.55	4.84	25.83	0.23	0.36
			Comm. A	—	—	—	26.84	0.51	0.44
			Comm. B	—	—	—	25.84	0.23	0.30
			Comm. C	—	—	—	32.05	0.45	0.40
Baghouse	830201	aggreg.	Queen's	0.08	81.40	3.07	—	—	—
			EMR-ERL	0.57	81.45	3.05	17.09	0.17	0.70
			Comm. A	—	—	—	17.27	0.33	0.20
			Comm. B	—	—	—	16.28	0.31	0.18
			Comm. C	—	—	—	16.58	0.24	0.26
Recycle	830201	1700	Comm. A	—	—	—	24.06	0.27	0.29
			—	—	—	—	28.78	0.46	0.29
	830201	2130	Comm. A	—	—	—	21.72	0.24	0.29
			—	—	—	—	26.64	0.18	0.30
	830330	0000-0100	Comm. A	—	—	—	14.22	0.19	0.25
			—	—	—	—	14.74	0.21	0.17
Bed product	830202	0800-1000	Comm. A	—	—	—	0.93	0.21	0.06
			—	—	—	—	1.00	0.27	0.00
	830215	0300-430	Comm. A	—	—	—	0.76	0.18	0.09
			—	—	—	—	0.67	0.13	0.07
	830225	0600-0800	Comm. A	—	—	—	0.49	0.26	0.10
							0.45	0.36	0.06

We had our commercial analyst A repeat some of his analyses a few weeks after the first, with the results shown in Table 16.1. The "standard errors", measures of repeatability, computed from these data, are summarized in Table 16.2.

The results from analyst C were received last, after all computations in this report had been done and too late to influence them unless a serious discrepancy were clearly indicated, making revision mandatory. We note, though, that his result for the carbon content of Minto coal agrees well with the finding of EMR-ERL. We know that this analyst and the one at EMR-ERL used their instruments in optimum fashion and performed their analyses with the greatest care. It appears rather likely that the carbon content of the Minto coal is in fact about $66.5 \pm 0.2\%$ as indicated by these analysts's results, rather than the median value of 68.3% which we have used in our calculations. The difference would slightly shift the overall carbon balance, Fig. 8.16, very slightly improving it.

Proximate analyses of the product solids were performed in our laboratory. These analyses yielded essentially only ash and fixed combustible matter (FCM), the samples being dry as they came from the process and almost devoid of volatile combustible matter (VCM). The total combustible matter (CM, the sum of FCM and VCM) should agree quite closely with the sum of carbon, hydrogen and nitrogen if we assume that sulphur present in the solids is largely in the sulphate form. Sulphate sulphur is assumed to be inert in proximate analysis, remaining in the ash, whereas any organic sulphur and sulphide (pyritic) sulphur are oxidized and contribute to the apparent combustible matter. The comparison between mass fraction of C, H and N (as determined by us) in the bed, cyclone and baghouse products is shown in Figs. 16.1--16.3. For the bed product, $w_{CHN} \equiv w_C + w_H + w_N$ is a little greater than w_{CM} . It appears that this is largely attributable to overestimation of hydrogen and possibly nitrogen (Fig. 9.1, for example, shows unreasonably high H/C ratios for

Table 16.2. Standard errors in the determination of C, H and N, as indicated by the results from analyst "A" in Table 16.1 and by the analyses in triplicate reported by analyst C. The quantity W_i is the mass ratio (mass of element)/(mass of dry solid), and s_i is the standard error in determining W_i .

Element i	Range of W_i	Error s_i	
		Analyst A	Analyst B
C	0—0.05	0.006 W_C	0.0002 + 0.0014 W_C
	0—0.25	0.21 $W_C^{1.31}$	0.0002 + 0.002 $W_C^{1.25}$
H	0—0.005	0.0005	0.0001
N	0—0.003	0.0003	0.0001

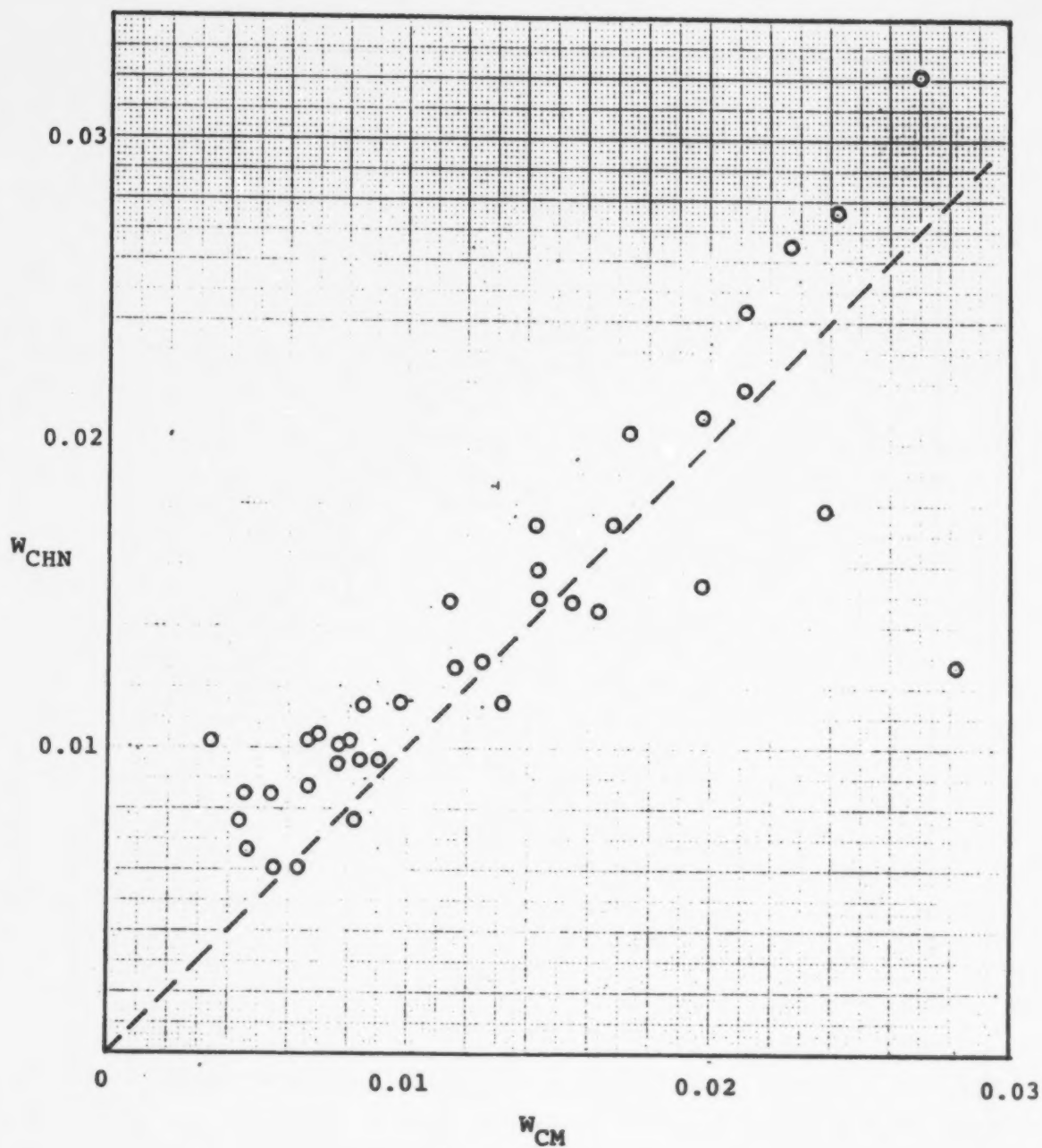


Fig. 16.1. Mass fraction C, H, N in the bed product vs. mass fraction combustible matter.

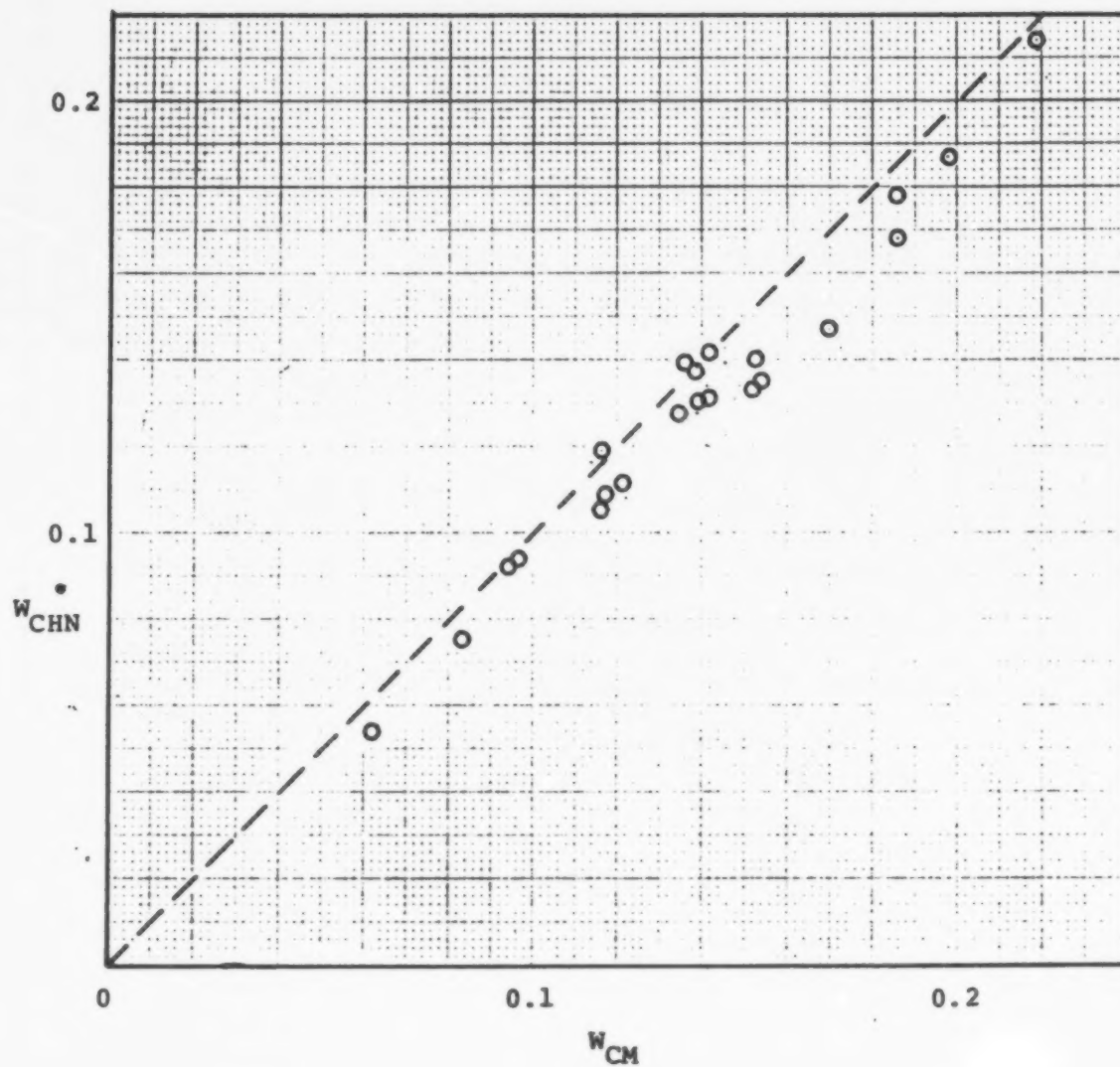


Fig. 16.2. Mass fraction C, H, N in the baghouse product vs. mass fraction combustible matter.

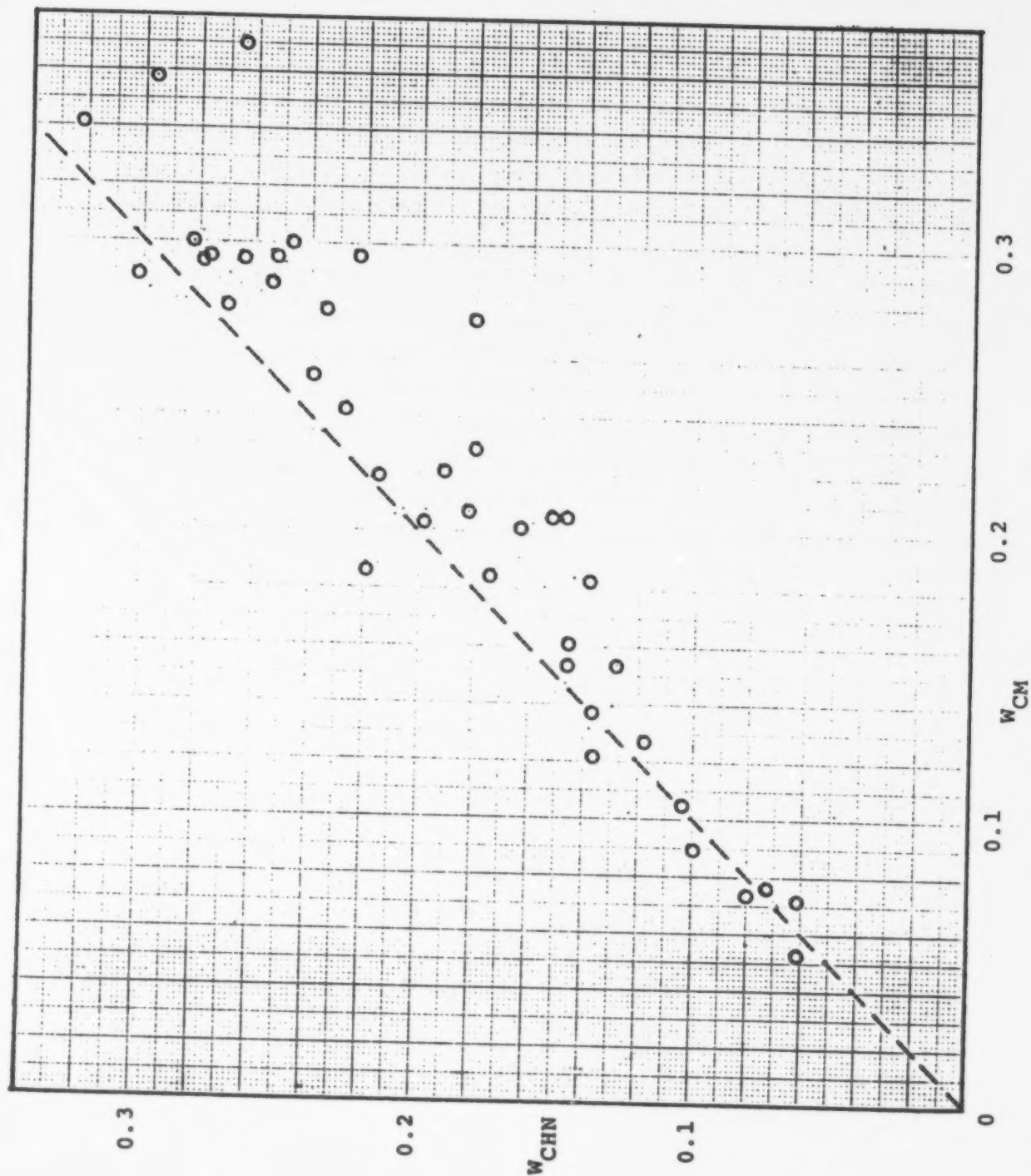


Fig. 16.3. Mass fraction C, H, N in the bed product vs. mass fraction combustible matter.

the bed product). Figure 16.4 shows that the carbon content w_C was generally a little less than w_{CM} , as expected. For the cyclone product, w_{CHN} agrees quite closely with w_{CM} up to $w_{CM} = 0.15$ but tends to fall off at larger w_{CM} . The reason for this is uncertain: it could be because carbon was underdetermined in the CHN analysis or/and because a substantial portion of the sulphur here was sulphide (pyritic) sulphur which boosted the combustible matter. The total sulphur content of the cyclone product was typically under 5 %, so it appears that the sulphide sulphur factor can only have been partly responsible. For the baghouse product the agreement between w_{CHN} and w_{CM} is quite close, and there is much less scatter than in the case of the cyclone product. However, it is apparent that w_{CHN} tended systematically to be slightly less than w_{CM} . The likely reasons for this are the same as for the cyclone product: underdetermination of carbon and/or presence of significant pyritic sulphur.

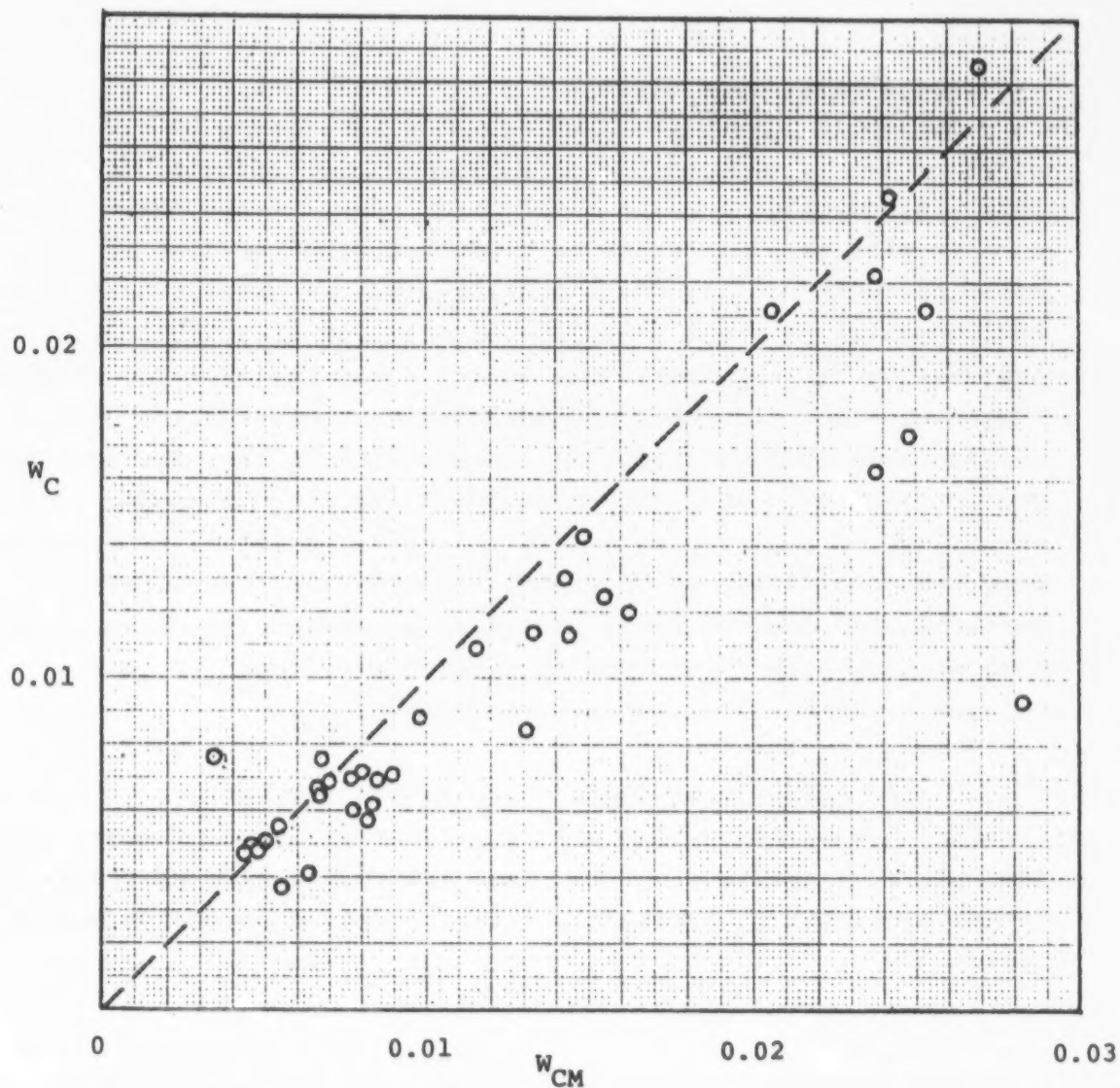


Fig. 16.4. Mass fraction carbon in the bed product vs. mass fraction combustible matter.

17. OTHER CHARACTERISTICS OF THE SOLID PRODUCTS

17.1 Introduction

It will be of interest to conduct further studies of the solid products, to see what handling problems they present, what environmental problems they pose if disposed of as solid wastes, and what useful applications they may have. As expected because of the substantial lime (CaO) content, the solids all evolve considerable heat when wetted or mixed with water. The cyclone product in many cases, when mixed with the right proportion of water, set to a hard plaster-like material, as might be expected from the combination of good CaSO_4 content and a favourable particle size distribution. Any work to be done in these areas, however, will likely be left to others to perform; it is certainly outside the scope of the present study.

Characteristics of the product solids that affect fluidization or combustion are of present interest, though. In the following two sections we cover some properties that were investigated but that have not arisen anywhere in the preceding discussion.

17.2 Bulk density

Bulk density of the solids was measured by very slowly pouring the material tested into a container of known volume, filling it without any jarring or vibration or any other disturbance that might cause consolidation, and then weighing it. The bulk density so measured is given by $\rho_b = m/V$, where m is the sample mass and V is the container volume. The avoidance--as far as possible--of consolidation should result in a value of bulk density quite closely approximating that obtained when the gas flow to a fluidized bed is very gradually reduced until the bed subsides and becomes "fixed" or "settled". Such values of bulk density can be used to estimate the settled bed height H_s when the holdup per unit base area is known.

The bulk density of the bed product was, unsurprisingly, found to be correlated with the Ca/S ratio, the factor that most strongly affects the conversion of CaO to CaSO₄. For operation with Minto coal

$$\rho_b = 1450 - 83 \text{ Ca/S}, \quad (17.2.1)$$

kg/m³. The bulk density of the cyclone product was 920 ± 70 kg/m³, and showed no simple trend with composition. The density of the baghouse product was determined only in one run, and was then 540 kg/m³.

The bulk densities of the product solids from operation with Devco coal were all lower than for Minto. The bed product (at the same Ca/S ratio) and the cyclone product were about 0.8 times as dense. The baghouse product had a density of only about 360 kg/m³.

Measured bulk densities of the bed and cyclone products were reproducible within 1--2 % for a given material. The baghouse product, however, was relatively loose and compressible, and reproducibility was only within 5--6 %.

17.3 The gross calorific value

The gross calorific value of a number of samples of the product solids was determined with a Parr combustion calorimeter. Since the samples were dry and had only small hydrogen content, the difference between gross and net calorific values was negligible.

The calorific values found for the cyclone and baghouse products are shown as a function of the content of total combustible matter (CM) in Fig. 17.1. The calorific value of the bed product could not be measured reliably because the combustible content was too low. Linear regression analysis of

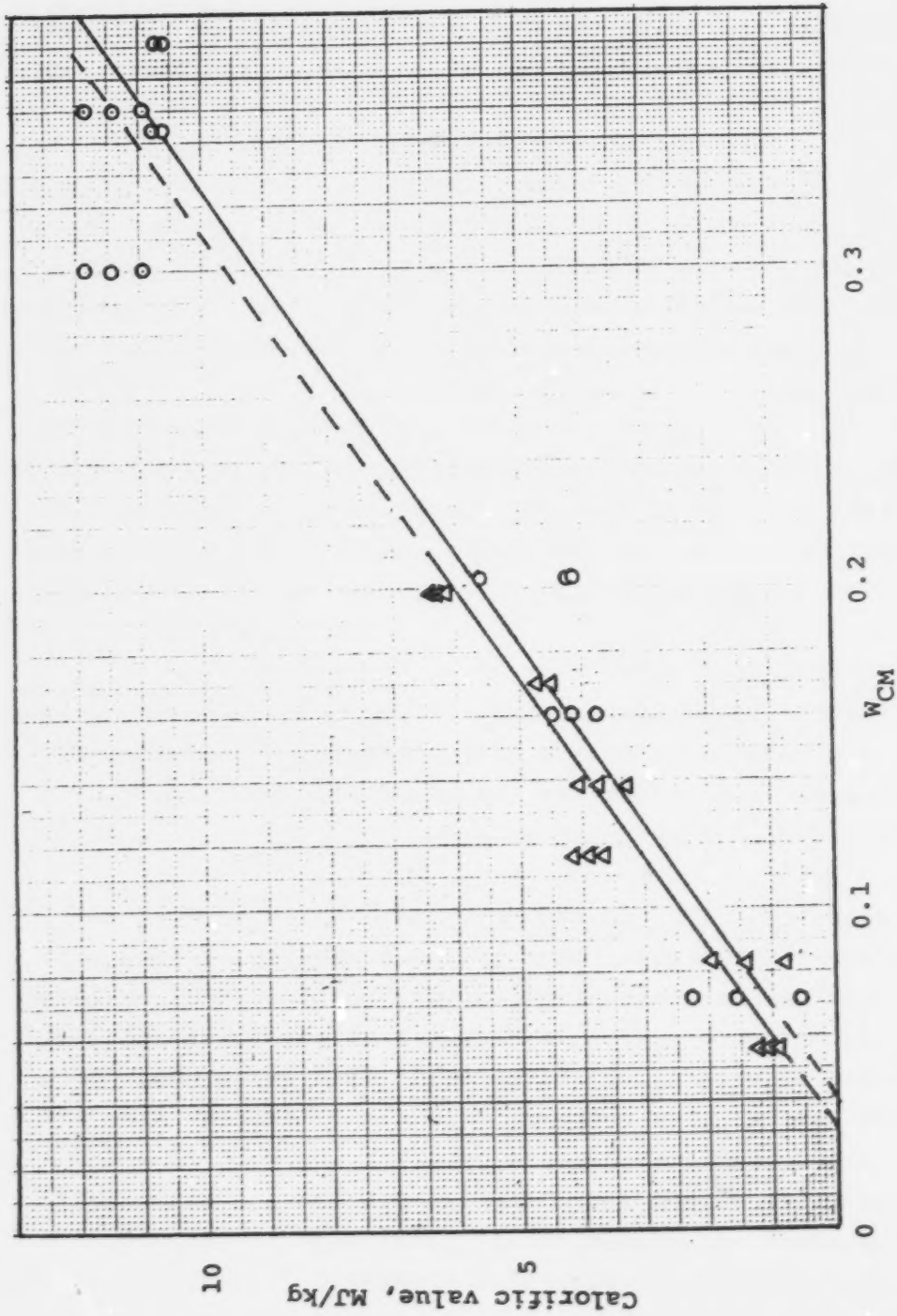


Fig. 17.1. Calorific value of cyclone and baghouse products vs. mass fraction combustible matter. Product: O, cyclone; Δ , baghouse.

the data in Fig. 17.1 gives significantly off-zero intercepts at zero combustible content. The slope, however, is virtually the same for both sets of data and indicates a calorific value of 35 MJ per kilogram of combustible matter. Since the enthalpy of combustion of carbon is 32.76 MJ/kg, this value is quite reasonable (a slightly greater value than for pure carbon is expected because of the small but not insignificant hydrogen content of the coal char).

18. CONCLUSION

18.1 Introduction

In this, the last section of this report, a discussion is given relating our work to that of others who have investigated the same coals and limestone. Comparisons with work on other coals and other limestones would open a very long inquiry; these will therefore be saved for later, more specialized publications.

The discussion is followed by a summary of principal findings, a list of practical recommendations, and a list of points for further study.

18.2 Discussion of results

18.2.1 Attaining steady state operation

The full achievement of steady state operation seems to have been treated fairly casually in many past investigations. It has often been assumed that steadying of the bed temperature and the exhaust gas composition are sufficient indications of effective entry into the steady state. These may be adequate criteria for these variables alone, but not necessarily for others, and particularly for the bed and bed-product compositions. As an extreme illustration, suppose that a bed has been operated with 3 mm dia limestone particles which, because of pore-plugging, cannot be more than 20 % sulphated. The limestone feed is then changed to one of 0.5 mm particles which can be 50 % sulphated. The gas composition may respond quite quickly to this change because the time constant for the sulphation of the 0.5 mm particles may be fairly small and so a virtual steady state in the characteristics of reactive limestone particles present in the bed may be attained rather quickly. However, the analysis of the bed particles will reflect the new situation only when nearly all of the original particles have been flushed out. When the bed consists 50 % by mass of the original 3 mm particles, and 50

% of the new 0.5 mm particles, the chemical composition of the bed product will have gone only about half-way towards the new steady-state composition. When only 10 % of the original material is left, however, the difference may be practically negligible. We have thus given the sure and sufficient criterion for steady state operation in terms of the residence time of bed product: steady state can reasonably be expected to be closely approximated under all circumstances when $t > 2 \tau_b$. This expectation is supported by the experimental results. When the starting bed material is quite like that in the eventual steady state, the time effectively required may be considerably less, but it appears that a virtual steady state can never be presumed to occur sooner than $t = \tau_p$. There are, of course, other factors besides reaction time involved; for example, the fluidization quality in a bed of 3 mm particles is likely to be significantly different from that in the case of 0.5 mm particles, and this may affect the steady state values of some combustion and sulphation performance characteristics.

18.2.2 Sulphation of limestone

Burning of Minto coal by fluidized bed combustion with addition of Havelock limestone for sulphur capture has been studied before, by Stover, Anthony, Desai and Friedrich (1981). They used a 240 mm I.D. tubular combustor, 3 m high from distributor to top of freeboard, with wall cooling of the bed but with no in-bed tubes. The expanded bed depth was about 1 m. The mass-median limestone particle diameter $\hat{D}_p$ was 1.15 mm, and that of the coal was 0.88 mm. Bed temperature was about 850 °C, fluidizing velocity about 2.2 m/s, and excess air 6--30 %. Underbed pneumatic feeding of the solids was used. No recycle was employed. With all these conditions virtually constant, variation of the Ca/S ratio produced the data on sulphur capture compared with ours in Figs. 11.3, 11.5 and 11.6. These results are lower than might be expected from ours by about 10 percentage points. It should be noted that the particle size distributions of both coal and limestone were considerably broader than any of

ours and this may have had some effect; however, it seems unlikely that it could explain more than a minor part of the difference in sulphur capture. The mean coal particle size, 0.88 mm, was very much smaller than our smallest, 6.6 mm, and this may have had some influence. Possibly important is the fact that bed material was removed from the bottom of the bed through a drain pipe in the middle of the distributor plate. Since coal and limestone were fed not far away, from a port in the combustor wall 100 mm above the distributor plate, there is a distinct likelihood of some short-circuiting between the feed and drain points, and this would certainly decrease the sulphur capture. The drainage of bed material was only carried out periodically, though, rather than continuously, and this would considerably diminish the effect.

A bench-scale study of Havelock limestone as an SO_2 solvent for use in FBC has been made by Hamer (1983). Discussion of his results will be reserved for another report in which our own bench-scale results will be presented (the Ph.D. thesis research of Mr. M. Couturier).

18.2.3 Carbon loss

Stover, et al. (1981) also carried out trials in which carbon burnout was studied. The same Minto coal, $\hat{D}_p = 0.88$ mm, as in the sulphation trials was employed. However, no limestone was used. Instead, the bed was charged with a batch of silica sand, $\hat{D}_p = 1.13$ mm, no sand was fed thereafter, and no bed product was removed. Thus all solids removal was by elutriation. Presumably the fine coal-ash was all elutriated while rock particles above a certain size, accompanying the coal, accumulated in the bed. The carbon loss by the elutriation was found to be correlated by

$$\dot{m}_{C,e} = 0.00205 \dot{m}_{C,f} (v/\phi_a) \exp(3500/T_b) \quad (18.2.1)$$

where v is in m/s. This expression agrees in general form with

the present correlation for overhead carbon loss, (8.2.6), but the coefficients (or exponents) are all different in value: 1 instead of 0.86 on v , - 1 instead of - 0.5 on ϕ_a , and 3500 instead of 5200 on $1/T_b$. The effects of limestone particle diameter $\hat{D}_p$ and recycle fraction r , of course, are missing because these were not varied. In order to obtain comparisons with our results, we have introduced an effect of bed-particle diameter $\hat{D}_p$ into the correlation of Stover, et al., assuming (as with our results, Sect. 8.2.10) that the exponent on $\hat{D}_p$ is - 1/1.163 times that on v . Further, for the effect of recycle fraction we have inserted our factor of $\exp(-0.36 r)$. The modified correlation then reads

$$\dot{m}_{C,o} = C_o \dot{m}_{C,f} v \hat{D}_p^{-0.86} \phi_a^{-1} \exp(3500/T_b) \exp(-0.36 r), \quad (18.2.2)$$

where in the work of Stover, et al., $\dot{m}_{C,o} = \dot{m}_{C,e}$, $\hat{D}_p = 1.13$ mm, and $r = 0$. Point-by-point values of C_o so calculated from their data and ours are shown as a function of bed temperature in Fig. 18.1. According to this representation, it appears that they typically found about 40 % more carbon loss than we. They covered a larger temperature range and their data suggest a maximum in the carbon loss around 850 °C.

The alternative approach to making a comparison is to calculate C_o for both data sets as defined by our correlation, (8.2.7). The results of this are shown in Fig. 18.2. As expected, the data of Stover, et al. (1981) show slightly more scatter in this form and ours show less. The maximum in their results (or is it here the start of a plateau?) moves to higher temperatures, perhaps around 900 °C. Their results remain around 40 % above ours.

Statistically (see Table 8.2, for example) it appears that the differences between us and Stover, et al. (1981) in the

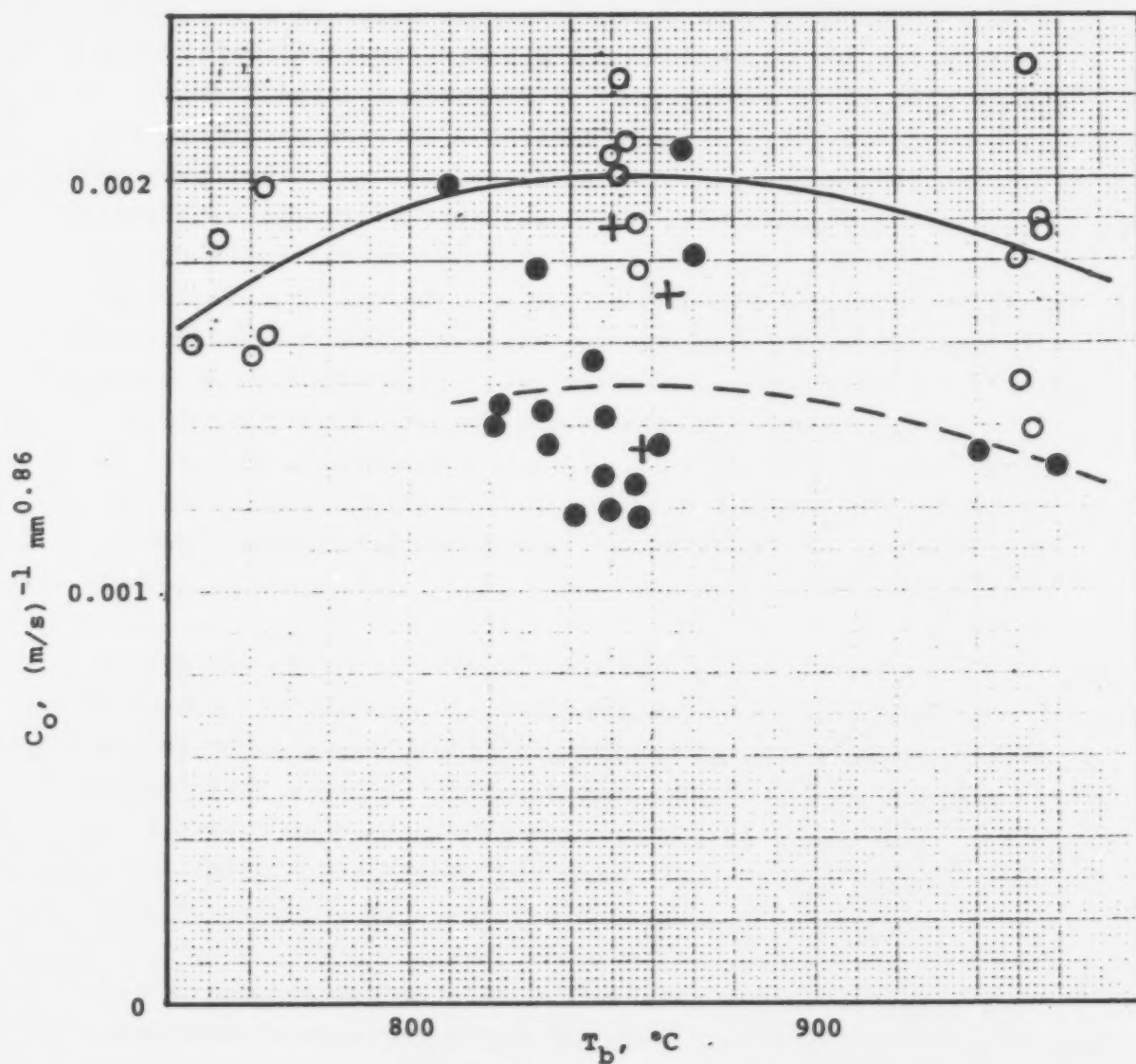


Fig. 18.1. The coefficient C_o in the equation (18.2.2) for overhead carbon loss based on the correlation of Stover, et al. (1981), computed point-by-point, vs. bed temperature. Coals: $\circ \bullet$, Minto; $+$, Devco. Sources: $\circ$, Stover, et al. (1981); $\blacksquare +$, present data.

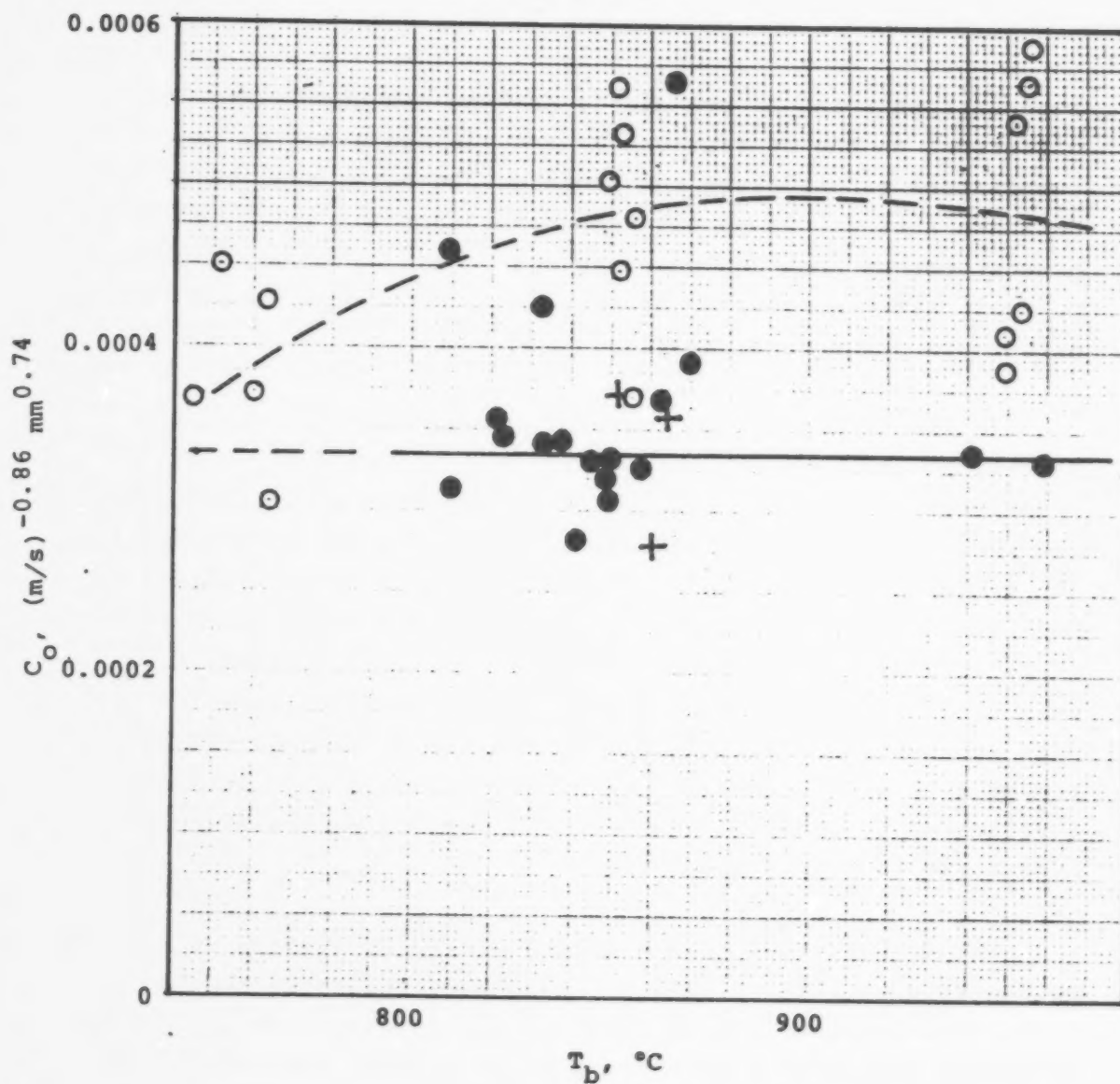


Fig. 18.2. The coefficient C_o in the present correlation (8.2.6) for overhead carbon loss, computed point-by-point. Coals: $\circ \bullet$, Minto; $+$, Devco. Sources: $\circ$, Stover, et al. (1981); $\bullet +$, present data.

exponents on v and ϕ_a are small, perhaps insignificant. The difference in the coefficient of $1/T_b$ is explained by the temperature ranges; over the range of our data, Fig. 18.2, their data are fitted well by our temperature coefficient. The 40 % greater carbon loss indicated by their data, however, appears to be real, and constitutes the one undoubtedly significant difference between our results.

In the matter of the temperature effect, the results of Stover, et al. suggest that the assumed Arrhenius-type temperature dependency may not be a very good approximation over the whole temperature range 750--950 °C. In the range 750--850 °C, a coefficient for $1/T_b$ of about 2000 K fits their data, at 850--950 °C it is about 5100 K, and over the whole range 750--950 °C it is 3500 K. It appears though, that for 810--960 °C, if not down to 750 °C, our value of 5200 K is reasonably good for our data and theirs. This is roughly equivalent to a power-law temperature effect of $T_b^{4.5}$, a strong enough effect so that attribution to chemical kinetics (as implied by the Arrhenius-type form) is fairly plausible.

The 40 % greater carbon loss observed by Stover, et al. may be due to a variety of factors which were different in their work: the introduction of sand instead of limestone as the bed material, the absence of a bed product stream, the absence of in-bed tubes (affecting quality of fluidization and thus perhaps the elutriation), and the very much smaller mean coal particle size and broader size distribution. However, perhaps the question that most needs answering is why the difference in carbon loss was not greater. In particular, why does coal particle size not have a greater effect? We used Minto coal of mean particle diameter $\hat{D}_p$ from 6.6 mm to 19.3 mm and found no effect of coal particle diameter. Here Stover, et al. used Minto coal of mean diameter 0.88 mm, and the carbon loss was only 40 % higher. It may be asked, what mechanism of carbon particle generation could lead to overhead carbon losses that are almost

or quite independent of coal particle size over such a broad range? The only obvious and simple hypothesis that has occurred to us is that the coal contains small, relatively unreactive carbonaceous bodies, typically under 1 mm dia., which are freed by the combustion and disintegration of the material around them as a coal particle burns. The quantity of these bodies liberated depends essentially on their concentration in the raw coal, and not on the diameter of the coal particles as fed, provided that the said bodies are small relative to the coal particles in which they are embedded (i.e., their number and size have been negligibly affected by the preparatory crushing of the coal). It is postulated that these slow-burning bodies account for the bulk of the carbon elutriated. This is a very different hypothesis from that which claims that each coal particle as fed burns down with negligible physical shedding of carbon, and the small carbonaceous particles present in a bed are simply the shrunken residues of the original coal particles.

The dependency of carbon burnout on operating conditions evidently varies substantially between coals of different types. Stover, et al. (1981) carried out trials like those on the Minto coal for a Western sub-bituminous coal (Highvale) and found the carbon loss to have no significant dependence on fluidizing velocity and mixture strength, but to be dependent on temperature:

$$\dot{m}_{C,e} = 5.6 \times 10^{-6} \dot{m}_{C,f} \exp(8400/T_b) \quad (18.2.3)$$

This indicates a loss of $0.010 \dot{m}_{C,f}$ at 850 °C, very much smaller than for the Minto. The correlation (18.2.3) is very weak, though, the correlation coefficient being only 0.53. The correlation actually considerably misrepresents the observed temperature dependence. Figure 18.3 shows the relative carbon loss $\dot{m}_{C,e}/\dot{m}_{C,f}$ directly as a function of temperature, with different data symbols for different levels of fluidizing velocity. It appears that at the higher temperatures, say $T_b >$

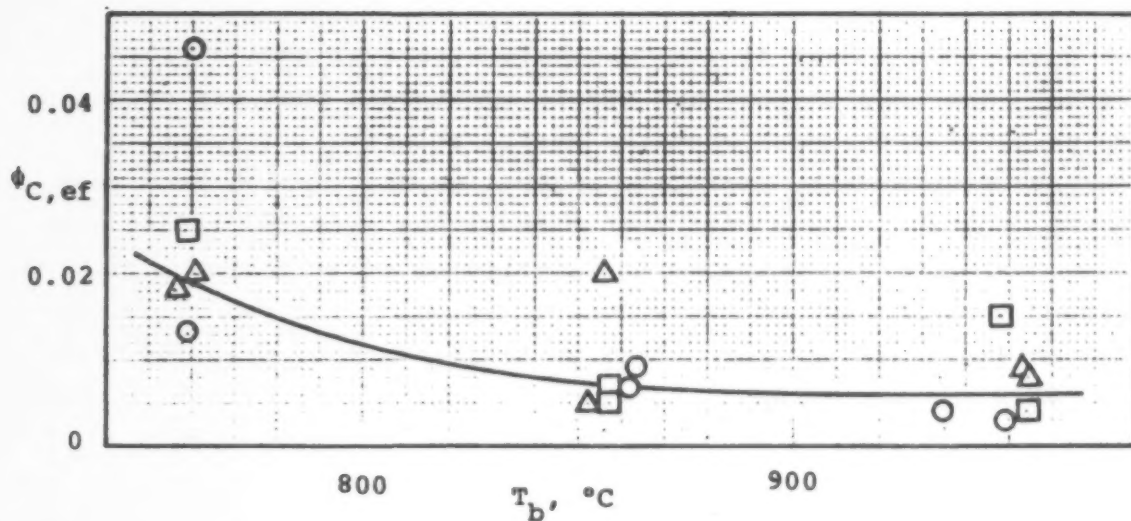


Fig. 18.3. Results of Stover, et al. (1981) for carbon loss in burning Highvale coal: flux ratio of carbon lost to carbon fed in fuel, vs. bed temperature. Levels of fluidizing velocity, m/s: $\circ$, 1.16--1.29; $\square$, 2.16--2.48; $\triangle$, 3.69--4.24.

840 °C, the temperature effect, if any, is quite small. A plausible explanation of this behaviour according to our simple hypothesis is that the Highvale coal contains a low mass concentration of relatively very unreactive carbonaceous bodies. These may be so unreactive simply because they are chemically very refractory, or it may be that, though correspondingly very small in number, they are large in particle size (thus low in surface/volume ratio) and therefore require much longer to burn. In any case, at $T_b > 840$ °C the carbon present in the bed is reduced to a minimum consisting of these bodies which are continually released by burning of the coal but which tend to be elutriated before they undergo substantial burning.

The only other item of evidence from other studies of the FBC combustion performance of the coals here of interest has to do with the effect of recycle. Unpublished results (private communication) from trials with Devco coal on the pilot plant at the Canadian Combustion Research Laboratory (Energy Research Laboratories, CANMET, Energy Mines and Resources Canada, Ottawa) indicate that recycle of cycloned dust can give as much as 50 % reduction in carbon loss. This agrees roughly with our indication of a limit of 30 %. The combustor of the CCRL pilot plant is the twin of ours, built from the same plans. Thus the comparison of results is uncomplicated.

18.2.4 Apparatus dependence of results

Although there is evidence that our results are of fairly general validity, a not insignificant element, in some ways at least, of apparatus dependence should be remembered:

- o the freeboard height was evidently not so great that reactions went virtually to equilibrium. A change of freeboard height would thus produce somewhat different results
- o our cyclone operated with a certain efficiency. Another cyclone might produce somewhat different results. The

overall system performance may thus be affected by the cyclone characteristics.

These examples are enough to show that a certain problem exists. Many others, mostly less serious, are easily identified.

18.3 Summary of findings

Following is a list of principal observations and conclusions resulting from the present study:

1. The time required to effectively reach steady state ranges from as little as one mean solids residence time $\tau_p = m_b / \dot{m}_p$ for stabilization of bed temperature and exhaust gas composition to $2 \tau_b = m_b / \dot{m}_b$ for complete stabilization of all significant aspects, including bed solids composition.

2. At all times, there are slow fluctuations in temperatures, compositions and flowrates due to unavoidable variations of coal and limestone feedrate, since these feedrates can never be made absolutely steady. The fluctuations may also be partly due to fundamental reactor instabilities. In any case, the fluctuations are in many ways highly revealing about the transient response of the system. A most remarkable feature is the virtually perfect positive correlation between the exhaust-gas concentrations of NO_x and O_2 . This shows again that the conversion of fuel nitrogen to NO_x in FBC varies most strongly with the excess air level, in the instantaneous sense as well as in the time average.

3. Even with underbed feeding of fuel, the mode used in all the present trials, freeboard reaction is significant. As expected, the freeboard burning produces drops in O_2 and rises in CO_2 levels. The SO_2 level sometimes falls and sometimes rises, or does both, passing through a maximum. This complex behaviour evidently results from the pressure of several competing processes, such as burning of char and SO_2 capture or loss by particulates. The NO_x level either remains steady or drops,

sometimes by as much as 50 %.

4. Recycle of cycloned solids can decrease carbon loss by as much as 30 %. It also has many other effects. It slightly increases the heat transfer coefficient to in-bed heat transfer surfaces. It strongly affects the division of product solids between bed product and elutriation products (the cyclone and baghouse products). The baghouse loading typically triples under strong recycle, compared to no recycle at all. The effect that can be most dramatic is a negative one: too high a recycle level can enormously increase the dust burdens on the cyclone separator and the recycle system, to the point of plugging.

5. Carbon burnup efficiencies were in the range 87--98 %. High burnup is promoted by the expected factors such as adequate excess air, high bed temperature, low fluidizing velocity, and reasonably strong recycle of cyclone product. Less expectedly, burnup increases with increasing limestone particle size, and it is apparently independent of coal particle size down to around $\hat{D}_p = 1$ mm (and possibly less). An empirical model has been developed which quantifies these effects:

$$\dot{m}_{C,o} = 0.000335 \dot{m}_{C,f} v^{0.86} \hat{D}_p^{-0.74} \phi_a^{-0.5} \cdot \exp(-0.36 r) \exp(5200/T_b) \quad (8.2.6)$$

A mechanism has also been suggested which physically explains the facts. Briefly, the coal, as it burns, continually sheds particles of relatively unreactive carbonaceous matter. These particles make up the bulk of the elutriated carbon, and their production is essentially independent of the coal particle size (provided that the coal has not been crushed or ground too finely). The cyclone product is usually the main carrier of lost carbon, and the bed product is minor. At high recycle levels, the baghouse product becomes an important carrier.

6. The Havelock limestone gives quite good sulphur capture when supplied in mean particle sizes under 1 mm. The most important factor then affecting the level of capture is the Ca/S ratio itself. The following expression correlates the results reasonably well at $\text{Ca/S} > 1.5$:

$$\eta_{\text{SO}_2} = 100 (1 - 2.6 \exp(-1.33 \text{ Ca/S})) \quad (18.2.1)$$

No effect of bed temperature was discernable in the range 800--880 °C. There were indications from the product solids analysis, however, of decreasing sulphur capture at temperatures over 900 °C. It appears that sulphur capture was slightly better with the Devco coal. This, however, is likely to be due to a difference in pyritic sulphur bodies and proportions between Devco and Minto coals, since it is unlikely that the coal type affected the reaction between SO_2 (once formed) and calcined limestone. There were no other discernable differences in FBC performance between the coals.

8. The bed, cyclone and baghouse products generally tend to be quite distinctive in appearance and chemical composition as well as in particle size. The bed product is the major carrier of calcined and sulphated limestone, and thus of Ca, Mg and S. The fine (non-stony) portion of the coal ash, consisting mainly of very small particles, is concentrated in the baghouse product which is, accordingly, rich in Fe, Al and Si. The high iron content is visible in the distinctly dark red colour of the material, like a darkened hematite. The high concentration of fine coal ash in the baghouse product may have environmental implications in respect to heavy metal pollution.

9. Bed-side heat transfer coefficients on in-bed cooling tubes are high, 200--350 W/m² K. They appear to be only weakly affected by particle size of the bed material and by fluidizing velocity.

18.4 Practical recommendations

The following recommendations are offered for practical operation:

1. For maximum effectiveness for sulphur capture, the Havelock limestone should be used in mean diameters $\hat{D}_p$ of 0.5--1.0 mm. Particles that are much smaller are elutriated too readily. Larger ones result in substantially increased limestone requirements for a given level of sulphur capture. Particles in the recommended range should give a good quality of fluidization at fluidizing velocities of 1--4 m/s.
2. There is evidence that increasing limestone particle size decreases overhead carbon losses. This is a reason for favouring the largest limestone particle size that gives good sulphur capture. A size around $\hat{D}_p = 1$ mm may be optimum.
3. Coal particle size, at least for Minto coal, seems to have negligible effect on sulphur capture and carbon loss over the experimental range $\hat{D}_p = 1$ --20 mm. We recommend, although we have not proven that it is optimum, the largest size we used, basically defined by passage through a 37.5 mm sieve. Coal particles of this size are small enough to move quite freely between in-bed tubes if they are reasonably spaced. However, note recommendations 4 and 5 below. Use of coal large particle size saves costs of finer crushing, of course, and reduces feeding problems associated with high moisture contents.
4. Screened coal of a given maximum particle size generally has with it stony matter (typically shale in the case of Minto and Devco) of similar maximum particle size. If these rock particles are too large to fluidize, they will naturally tend to form a growing stagnant layer at the bottom of the bed. It may then be necessary to make special provision to remove this material continuously or periodically without shutdown. If

underbed removal of bed product is used, the unfluidizable material (its angle of repose diminished by the fluidizing airflow) may drain satisfactorily. However, it may be necessary to resort to a sloping distributor plate or other devices to prevent stagnation.

5. Operation with Minto coal produced clinkering in some of our runs. To avoid clinkering, it is recommended that:

- o the combination of coal of large particle size (containing a substantial fraction of + 20 mm particles) with limestone of large particle size (containing a substantial fraction of + 2 mm particles) should be avoided, and certainly not combined with low levels of fluidizing velocity (say $v < 2 v_{mf}$). It appeared that the clinkering was generally associated with poor bed mixing and large coal particle size which evidently acted to produce hot spots.
- o the minimum distance between adjacent in-bed heat transfer tubes should be at least twice the diameter of the largest coal particles fed.

These suggestions are based on our experience with underbed feeding of coal and limestone. It appears likely that the tendency to clinker would be diminished by overbed feeding of coal.

6. There is negligible gain in recycling more than ten parts of cycloned dust for every one removed as product. It is recommended that when an efficient cyclone is used (such that little dust is passed on to the baghouse or other ultimate dust removal system), provision be made for removing about 1/10 of the cycloned dust as product while recycling the rest, rather than recycling everything. This will diminish the load on the cyclone and recycle system without significant loss of combustion efficiency or sulphur capture. It will also substantially

decrease the load on the baghouse. The cyclone product has interesting properties and may have practical applications such that its separation from the bed and baghouse products is advantageous. However, it may also happen that recycle cannot be economically justified, the gains being too small to justify the costs.

7. The fact that samples of bed material taken low in the bed contained significantly more carbon than the bed product, which was removed by overflow from the top of the bed, shows that a slightly greater carbon loss would have occurred had underbed removal of bed product been used. The effect would have been small because the bed product was in any case only a minor carrier of lost carbon. However, it is only reasonable to try, if possible, to locate the point of removal of bed product so as to minimize carbon loss. It is recommended that this removal point be placed as far as is practically possible from the feed points for coal and cyclone-dust recycle, to avoid short-circuiting. In overbed feeding, underbed removal is likely to be best. However, the removal point should be out of the path and proximity of the recycle stream. When coal of large particle size is used, so that removal of unfluidizable rock is a problem, then this may dictate underbed removal of bed product even when underbed feeding of the coal is used.

18.5 Points for improvement and for further investigation

Our work has not answered all the interesting and practical questions relating to the FBC burning of Minto and Devco coals, with limestone addition for sulphur capture. We would like to, and will if we can, pursue the following inquiries:

1. Overbed feeding, particularly of the coal. How does the system performance differ from that in underbed feeding? Coal of large particle size should be favoured since it can, for various reasons, be expected to perform better in this mode than coal of small particle size (less freeboard burning, better mixing into

the bed). The evidence presented herein shows that in underbed feeding, coal particle size (if not too fine) negligibly affects performance.

2. Operation at greater bed depths. The settled bed depth in the present work was around 0.5 m. We would like to explore operation at settled bed depths up to 1 m or more, to see how this affects carbon loss, sulphur capture, and freeboard processes. We already have evidence (Xu, Becker and Code 1983) that bed depth has very little effect on elutriation.

3. Kinetics of freeboard processes. We got much interesting information about freeboard gas composition in the present work. What we now want are some fast, accurate measurements of vertical concentration profiles, representing the conditions over a period of time short enough to be virtually unaffected by the slow fluctuations which have been remarked upon. These measurements must be complemented by equally accurate gas temperature measurements which can only be made with a suitably miniature aspirating thermocouple or similar device. The "freeboard temperatures" measured in the present work cannot be relied upon at all to represent gas temperatures; the thermocouples (as in most FBC pilot plants) were far too large and responded primarily to the radiant field around them.

4. Operation at very high recycle solids fluxes. In the present work we could not, in some cases, operate at total recycle of the cyclone product because the recycle system was then overburdened. We would like to give this question a final scrutiny, to see whether very high recycle fluxes produce any effects not already indicated by extrapolation to $r = 1$.

5. Residence time distribution in the bed for particles of different diameters (size-resolved RTD's). In the present work, only the mean residence times of total solid products (τ_p) and bed product (τ_b) were determined. It would be particularly

interesting to know the times available for burning of fine char particles before they are elutriated.

6. Phenomena and mechanism of the burning of relatively large particles of coal. This work would be done on our bench-scale system. Evidence relating to the mechanism of carbon loss by elutriation would be particularly interesting, to help explain the present empirical model and get experimental facts regarding the physics.

7. The existence and implications of multiple steady states. This would likely begin as a theoretical inquiry and might then be followed up by suitably designed experiments.

8. The optimization problem of selecting the best limestone particle size. Limestone particle size affects both sulphur capture and, as we have seen, combustion efficiency. Before proceeding with this, we will want to see the results of bench-scale studies that are currently nearing completion.

18.6 Closure

Fluidized bed combustion has been under intensive investigation for around two decades and many of the basic facts have been well established. The direction of progress now is towards refinement of the operation, testing of different coals and limestones, and higher development of mathematical modelling as a tool for prediction, design, optimization, and control. The present study contributes to these efforts in the following ways:

1. Knowledge has been advanced in understanding the effects of recycle and the mechanics of elutriation. The examination of the latter has been kept brief herein because a separate report is being issued (Xu, Becker and Code 1983).

2. The performance of two Canadian coals which are strong candidates for fluidized bed combustion, and of an appropriate

limestone from the same region, has been assessed.

3. Together with results from complementary bench scale experiments, the data are being used in basic investigations on calcination, sulphation, combustion, and mathematical modelling. As noted in the Introduction, the material coming out of these studies will be published in other reports, papers and theses.

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